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Flue Gas Desulfurization System Capabilities for Coal-Fired Steam Generators. Volume II. Technical Report

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FLUE GAS DESULFURIZATION SYSTEM CAPABILITIES FOR COAL-FIRED STEAM GENERATORS

Volume II. Technical Report

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The report discusses the availability of technology for reducing SO_2 emissions from coal-fired steam generators using flue gas desulfurization (FGD) systems. Foreign and domestic lime, limestone, double alkali, magnesium slurry, and Wellman-Lord FGD systems are described, and the design parameters and operating experiences are discussed. Steps that have been taken to achieve high system operability are discussed. Also, disposal of FGD system wastes is discussed briefly.

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by

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ABSTRACT

This study presents an assessment of flue gas desulfurization (FGD) systems with respect to both SO_2 removal potential and system operability. This assessment was made by reviewing the performance of operating systems and the features of new systems designed to alleviate previous problems. Five major FGD processes are described in detail. The impact of key design parameters on removal efficiency and process operability was investigated from a theoretical point of view and by using data from operating systems. In addition, the operating experience of major FGD installations was reviewed. Operating problems and their solutions were analyzed with respect to their impact on process operation. Research data were used to evaluate the status of process development and the potential for sustained system operation and high removal efficiencies.

This study shows that the major systems, comprising lime and limestone slurry, Wellman-Lord, magnesium oxide scrubbing, and double alkali processes, are capable of removing SO_2 with high efficiencies when applied to both high- and low-sulfur coal combustion facilities. Such results have been obtained in pilot, prototype, and full-scale systems. Though sustained operation at high SO_2 removal efficiencies has not been widely achieved, a basis for design of such systems has been developed. When required, new systems are being designed for 90-percent removal efficiency and an availability of 90-percent or greater.

An analysis of the effect of using spare scrubber modules is also presented and shows that large improvements in system availability can be achieved by installing a spare module.

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METRIC CONVERSION FACTORS

To convert English units	Multiply by	To obtain SI units
British thermal unit (Btu)	1054	Joule (j)
Cubic foot (ft^3)	0.0283	Cubic meter (m^3)
Degrees fahrenheit	5/9 (°F-32)	Degrees Celsius (C)
Foot	0.3048	Meter (m)
Gallon (U.S. Liquid)	0.0038	Cubic meter (m^3)
Gallon (U.S. Liquid)	3.7854	Liter (l)
Horsepower (hp)	746.0	Watt (w)
Inch	0.0254	Meter (m)
Inch	2.54	Centimeter (cm)
Inches of water	248.8	Pascal (pa)
Pound	0.4536	Kilogram (kg)
Ton, short	0.9072	Metric ton (kkg)

1.0 INTRODUCTION

The United States Environmental Protection Agency (EPA) has undertaken a program to review the current New Source Performance Standards (NSPS) limiting sulfur dioxide (SO_2) emissions from large coal-fired steam generators. The current emission regulation was promulgated in 1971, and EPA was recently mandated to review this regulation in light of new technology. EPA's Office of Air Quality Planning and Standards has prime responsibility for this review, which will form a basis for determining whether the current NSPS should be revised. To develop background information in support of this review, EPA has contracted with PEDCo Environmental, Inc., to evaluate the capabilities of flue gas desulfurization (FGD) techniques.

1.1 HISTORY OF FGD SYSTEMS

The concept of scrubbing flue gases from coal-fired utility boilers and other processes is not new. In 1926, the 125-MW coal-fired Battersea Power Station in London, England, was equipped with a spray-packed tower and a tail-end alkaline wash section. The process was more than 90-percent efficient in the removal of SO_2 and particulate from the combustion gas of coal with a sulfur content of 0.9 percent. The alkaline waters of the River Thames provided most of the absorptive alkali; about 20 percent was made up from lime addition. Manganese sulfate was added to the scrubber effluent to enhance the oxidation of calcium sulfite to sulfate. The solution from the scrubber passed to a settler and into an oxidizing tank before being discharged back into the Thames.

Later air and water pollution control programs, supported by Imperial Chemical Industries, Ltd., British Power Authority consultants, and the Howden Construction Co., led to the development of a closed-loop, lime-based system for removal of sulfur dioxide. This was installed at the Swansea power plant in 1935 and at the Fulham power plant in 1937. They operated successfully until the early period of World War II when they were shut down because the vapor plumes provided possible aerial guidance to enemy aircraft.

The next large-scale uses of lime/limestone scrubbing were in Russia and Japan. A scrubber installation has been operating in Russia since 1964, scrubbing sulfur dioxide from about 850 m^3/s (1.8 million acfm) of waste gas from an iron ore sintering plant. The first lime scrubber in Japan was on a large sulfuric acid plant in 1966.

In the United States, the Tennessee Valley Authority (TVA) conducted small-scale and limited pilot plant studies in the 1950's. The first major pilot plant work appears to have been that of Universal Oil Products (at a Wisconsin utility installation) beginning in 1965. A limestone slurry circulating through a mobile-bed scrubber treating 0.94 m^3/s (2000 acfm) gave good SO_2 removal.

In 1966, Combustion Engineering tested a system involving injection of limestone, followed by scrubbing, in a pilot unit [1.4 m^3/s (3000 acfm)] at a Detroit Edison power plant. At a stoichiometric limestone-to-SO_2 ratio of 1.1 to 1, the SO_2 removal was 98 percent. On the basis of this pilot plant work, the company offered the process to the utility industry, and five systems were installed. One installation was at the Union Electric Co. in St. Louis, Missouri (140 MW). Kansas Power and Light Co. installed one on a 125-MW boiler in Lawrence in 1968; another on a 400-MW unit at the same plant started up in 1971. Kansas City Power and Light has used the process on two boilers, one at 100 MW, and the other at 140 MW.

Because of major problems associated with dry limestone injection, these systems proved inadequate. The Union Electric installation has been abandoned, and the Kansas Power and Light systems are being replaced by a system in which a limestone slurry is introduced into the scrubber. The Kansas City Power and Light installations are being converted to lime scrubbing. Problems associated with limestone injection include plugging (especially of the boiler tubes), low sulfur dioxide absorption, and reduced particulate collection in the electrostatic precipitators.

1.2 APPLICATION OF FGD SYSTEMS

As of August 1977, there were 141 FGD systems with an equivalent electrical generating capacity of 54,840 MW planned or in operation in this country. Of these systems, 29 were operational (8,914 MW); 28 were under construction (11,810 MW); and 68 systems were planned (32,628 MW). An additional 16 installations (8,592 MW) are considering the use of FGD systems as well as other control strategies such as low sulfur content coal. Approximately 12 to 15 boilers (6000 MW) are planning to use FGD systems, but have not publicly released their plans.

1.3 REPORT CONTENTS

This study presents a brief overview of alternative SO_2 control techniques in Section 2 and their potential application for meeting alternative NSPS levels. Section 3 provides an evaluation of the current capabilities of the lime, limestone, double alkali, magnesium oxide, and Wellman Lord FGD processes and a brief review of several other FGD systems. Section 4 presents an overall analysis of FGD system performance and operability based on the information presented in Section 3.

Information for this study was obtained from the literature and from contacts with users and suppliers of FGD systems.

2.0 OVERVIEW OF ALTERNATIVE SO_2 EMISSION CONTROL SYSTEMS

To comply with current NSPS, operators of coal-fired steam generators with heat inputs greater than 260 GJ per hour (250 million Btu per hour) must limit SO_2 emissions to 0.52 g SO_2/MJ (1.2 lb $SO_2/10^6$ Btu) of heat input. Emissions can be controlled to the required level by one of the following methods: burning naturally occurring low-sulfur coal, lowering the sulfur content of the coal prior to combustion (coal cleaning), removing SO_2 from flue gases after combustion, or using fuel treatment processes and combustion systems that remove SO_2 prior to, or during combustion (coal gasification, coal liquefaction, and fluidized-bed combustion). Selection of a control technology for a new combustion source is determined by the degree of control required, process availability, and costs.

The following sections concern methods of controlling SO_2 emissions other than by flue gas desulfurization (FGD). These methods are discussed in terms of their possible application in new, coal-fired boilers used by utilities to meet the current NSPS level, as well as alternative levels, of 90 percent reduction of SO_2 emissions and 0.25 g SO_2/MJ (0.5 lb $SO_2/10^6$ Btu) heat input.*

2.1 LOW-SULFUR COAL

Reserves of coal in the United States suffice to meet projected, long-term demands for energy production, but only about 9 percent of the recoverable reserves are of sufficient quality to enable utilities to meet the current NSPS without fuel processing or emission controls (see Table 2-1). If a standard as stringent as 0.25 g SO_2/MJ (0.5 lb $SO_2/10^6$ Btu) were promulgated, less than 1 percent of the recoverable coal could be burned without some kind of flue gas treatment.

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* Alternative levels provided by EPA.

2.1.1 Definition

Coals can be categorized according to sulfur content in relation to SO_2 emissions, since approximately 90 percent of the sulfur is emitted as SO_2 when coal is burned. No definition of low-sulfur coal is universally accepted, however. Coal can be burned in compliance with the current Federal NSPS for boilers if it emits a maximum of 0.52 g SO_2/MJ (1.2 lb $SO_2/10^6$ Btu) gross heat value. Figure 2-1 relates fuel sulfur content and heat value required to meet the current NSPS SO_2 emission limit.[1]

Reported data on coal sulfur content represent nominal, long-term (annual) averages, and do not take into account short-term fluctuations. If a revised NSPS were issued based on a shorter averaging time, such as 24 hours, the required sulfur content of the coal would have to be much lower to satisfy the same emission limit. Accurate statistical data on sulfur variations in coal are required in order to predict the level of coal sulfur content needed to meet a standard that is based on averaging times of less than one year.

2.1.2 Coal Reserves

Data on low-sulfur coal reserves are not exact and are subject to estimation and interpretation. Estimated amounts of reserve base relate to coal found in seams that can be economically mined by conventional techniques. Not all

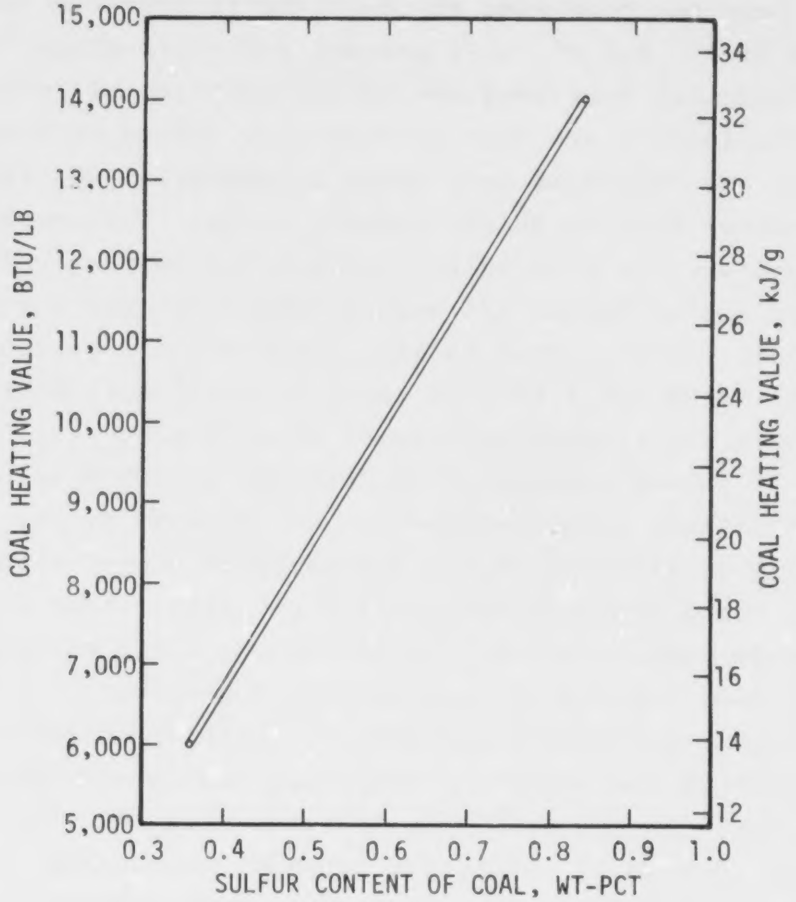

Figure 2-1. Plot of the maximum permissible sulfur content versus Btu content of coal required to meet current NSPS.

this coal can be recovered, however. Estimates must be reduced to allow for pillars for roof support, mining losses, overburden, etc. Traditionally, about 50 percent of the reserve base has been used to calculate recoverable reserves for deep mines, and 80 to 90 percent for strip mines.[5]

Recoverable coal reserves in the continental United States (Table 2-1) are derived from U.S. Bureau of Mines estimates and expressed on a basis of potential SO_2 emissions, rather than on sulfur content alone. This method takes into account both sulfur content and heating value and is a more useful way of estimating emissions from a particular coal. As mentioned earlier, only about 9 percent or 22.9 billion Mg (25.2 billion tons) of total coal reserves can achieve the current NSPS level of 0.52 g/MJ (1.2 lb/10^6 Btu) [0.7% sulfur content at 26,300 J/g (11,300 Btu/lb) of coal] without prior treatment. An estimate by the University of Illinois Center for Advanced Computation, however, based on older Bureau of Mines data, shows a total recoverable reserve of only 15 billion Mg (16.5 billion tons) of coal capable of meeting this standard.

Figure 2-2 displays the data in Table 2-1 graphically and shows that the amount of complying coal would decrease significantly if the NSPS were lowered. Its reduction to 0.25 g/MJ (0.5 lb/10^6 Btu), for example, would mean that less than 1 percent of the recoverable coal reserve could meet the standard without pretreatment or flue gas desulfurization. Available data on coal sulfur content below 0.43 g/MJ (1 lb/10^6 Btu) are sparse and the amount of coal with very low sulfur levels can be estimated only crudely. Current data, however, indicate that reserves are very limited.

Table 2-1. SUMMARY OF TOTAL RECOVERABLE COAL RESEPVES BY REGION AND EMISSION POTENTIAL - CONTINENTAL U.S. (millions of short tons)[a]

Coal Type/Region	Potential SO_2 Emission Rate, lb/10^6 Btu					
	≤ 1.2	1.201 - 1.6	1.6 - 2	2.01 - 3	>3	Total
Bituminous						
East	17,366	7,708	9,302	12,721	57,896	104,993
Midwest	925	1,571	2,017	2,189	290,811	97,513
West	4,167	7,103	43	331	1,022	12,666
	22,458	16,382	11,362	15,241	149,729	215,172
Subbituminous						
East						0
Midwest						0
West	0	19,219	14,870	4,754	0	38,843
Lignite						
East		(Quality of Reserves unknown)				1,026
Midwest	0	0	0	3,435	0	3,435
West	0	0	17,072	3,172	1,756	22,000
	0	0	17,072	6,607	1,756	26,461
Anthracite						
East	2,766	3,567	83	99	742	7,257
Midwest	0	0	0	17	17	34
West	9	0	0	0	0	9
	2,775	3,567	83	116	759	7,300
Bituminous	22,458	16,382	11,362	15,241	149,729	215,172
Subbituminous	0	19,219	14,870	4,754	0	38,843
Lignite	0	0	17,072	6,607	1,756	26,461
Anthracite	2,775	3,567	83	116	759	7,300
	25,233	39,168	43,387	26,718	152,244	287,776
% of Total	8.77	13.61	15.08	9.28	52.90	
Cumulative %	8.77	22.38	37.46	46.74	99.64	

[a] Multiply by 0.907 to convert to Mg

NOTE: Calculated from data in References 2, 3 and 4 by using the average heat value for each coal bed and assuming 95% conversion of sulfur to SO_2.

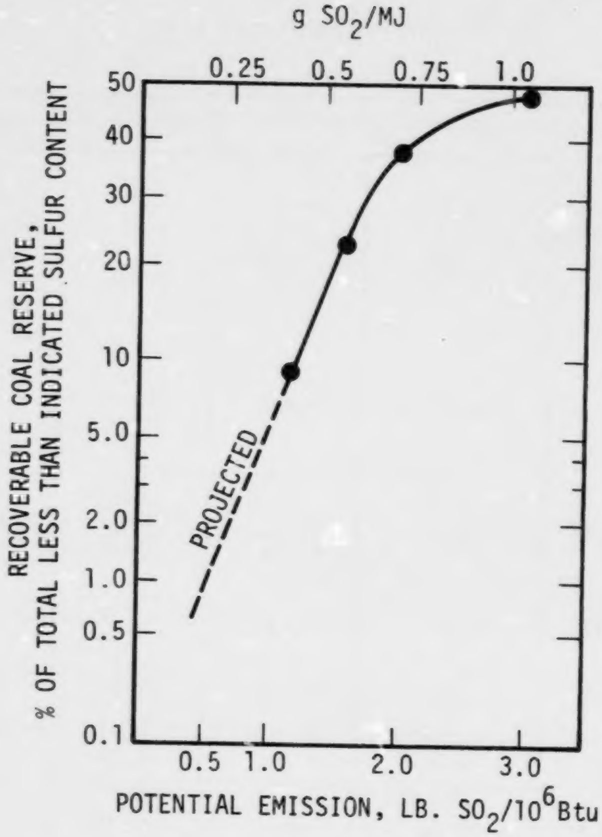

Figure 2-2. Recoverable coal reserve by sulfur content.

2.1.3 Other Factors Affecting Use of Low-Sulfur Coal

Other factors may determine whether coal that meets the current NSPS or alternative levels will be used. They include:

- Availability of mining and transportation equipment.
- Environmental impact from mining activity.
- New regulations affecting reclamation, safety, and utilization of regionally available coal.

2.1.3.1 Coal Mining and Transportation Equipment - The manufacture of mining equipment (especially large draglines for strip mining) depends largely on the willingness of mining companies to place firm orders well in advance of delivery, which in turn hinges on whether utilities will commit themselves to long-term contracts. The production of large draglines, however, has been increasing about 25 percent per year and this trend should continue for the next few years.[7] The equipment problem, therefore, may be affected more by a shortage of mining industry capital, than by insufficient manufacturing capacity. A possible shortage of hopper cars is related both to shortages of steel and castings, and to the unwillingness of railroads to invest money without strong projections on the demand for coal shipments. In the past, however, hopper car builders have been able to respond quickly to increases in demand, and it appears unlikely that hopper car delivery time will exceed that required to develop the coal mines. Also, utilities may buy hopper cars and enter into lease-back arrangements with the railroads.

2.1.3.2 Environmental Effects of Mining - Environmental effects of coal production result primarily from mining and transportation. Although specific effects depend on the location of the mine, some general tendencies are evident.

In strip mining, for example, topsoil and overburden are removed to reach the coal seam. The topsoil is either stored for later use, or used immediately to revegetate spoil piles or other disturbed areas. The ecological balance is always upset by mining operations. Even if revegetation efforts are successful, the previous balance of flora and fauna is never quite restored. In addition, removal of ground cover causes erosion and fugitive dust.

The environmental effects of underground mining include acidic mine drainage, unsightly piles of waste materials, land subsidence, and dust caused by coal handling and transportation.

Expansion of western coal production areas is hampered severely by lack of water, which is needed for revegetation, pipeline slurry transportation (if used), dust control on haul roads and in underground mines, coal cleaning operations, and for drinking.[8] Water shortage will undoubtedly be the major problem in the development of some coal fields.

2.1.3.3 Regulatory Factors Affecting Coal Mine Development - New State and Federal regulations requiring better reclamation of strip-mined areas and more stringent safety restrictions also affect further development of certain operating mines and construction of new ones.

The August 1977 amendments to the Clean Air Act (Sec. 125) will also affect the development of coal mines, because large coal-fired boilers may be prohibited from burning coal that is not mined locally or regionally.[9]

2.1.4 Conclusions

Limited amounts of naturally occurring coal are available to meet the current NSPS. Available data indicate that approximately 9 percent, or about 22.9 billion Mg (25.2 billion tons), of recoverable coal in the continental United States can be burned in compliance with the current NSPS SO_2 emission limit. If this limit is reduced to 0.25 g SO_2/MJ (0.5 lb $SO_2/10^6$ Btu), only a small portion (less than 1%) of the recoverable reserves will be able to meet the SO_2 limit.

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2. Ibid, p. 33-486.

3. Strippable Reserves of Bituminous Coal and Lignite in the U.S. Department of Interior, Bureau of Mines Information Circular 8531, 1971, p. 70-121.

4. Western Coal. Coal Age - Mid-April, 1973.

5. Op. cit. Ref. 1.

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9. Congressional Record - House, August 3, 1977 p. H8519.

2.2 PHYSICAL COAL CLEANING

Although coal beneficiation has been practiced for many years, especially in the metallurgical industry, application of this technique to meet SO_2 regulations, or to reduce the efficiency required of an FGD system, is a relatively new concept. Currently no coal cleaning plants are known to be producing coal that will meet current NSPS for boilers; but one plant designed for this (the Homer City Plant of Pennsylvania Electric Co.) is in the final stage of construction. It seems likely that coal cleaning techniques similar to those used in the metallurgical industry will find at least limited application as a control measure for complying with current SO_2 emission limits. However, as discussed in this section, any reduction in allowable SO_2 emissions will curtail this control option for new boilers.

The following subsections briefly describe several coal cleaning processes and assess the potential benefits and limitations of this control technology.

2.2.1 Coal Cleaning Processes

Coal is a complex mixture of organic and inorganic materials. Its physical and chemical properties are not homogeneous. Various materials, including sulfur compounds, are chemically bound within the coal structure or discretely dispersed through it. Although large inclusions can be liberated and removed by physical methods such as crushing, screening, grinding, and flotation, finely distributed mineral inclusions or chemically bound inorganic species usually do not respond to these methods.

The cleanability (lowering of sulfur and ash content) of coal depends on its size, the original sulfur content, the pyritic fraction (FeS_2), and the size distribution. A given coal can be cleaned to various sulfur levels by adjusting the size distribution and the effective separation density in the cleaning process. As the sulfur content of the coal is lowered, however, its heating value decreases because of carbon losses.

All physical coal cleaning processes include two basic operations: crushing, to a size that allows liberation of unwanted materials; and separation of these impurities, by exploiting the physical differences between them and the coal particles. Although some processes take advantage of differences in surface properties or magnetic properties, most utilize differences in density. The density of coal is about 1.3 g/ml, whereas the densities of most inorganic impurities found in it are much higher. The density of limestone for example, is 2.7 g/ml, and that of pyrite is 5.0 g/ml.

The washability characteristics of a given coal reserve can be determined by extensive core drilling. The samples are tested by float-sink techniques and analyzed. Because of the expense involved, these analyses usually yield only a limited amount of detailed washability data. When a coal bed is already being mined, washability characteristics are determined by actual production samples.

Four physical cleaning methods are now being used on a commercial scale: the dense-medium, hydraulic, froth flotation, and pneumatic processes. They are described briefly below.[1]

In the dense-medium process, a liquid of higher density than coal causes clean coal to float and the heavier impurities to sink. The principal dense-medium processes incorporate either vessels that rely on gravity, or cyclones that use centrifugal force to effect separation.

The principal hydraulic coal cleaning processes involve the use of jigs, launders, hydrocyclones, or wet concentrating tables. All these processes use water and movement or agitation to separate coal fractions.

Froth flotation exploits selective surface properties of certain coal solids. When a coal slurry is mixed with finely dispersed air bubbles, the bubbles preferentially attach to coal particles, rather than to ash particles, and bring them to the surface to be removed as a concentrate.

Most pneumatic cleaning methods use a pulsating upward air current to classify the coal into stratified layers. Ash and pyritic materials are withdrawn from pockets or wells at the bottom of the bed. Cleaned coal is taken from the top of the stratified coal column.

2.2.2 Potential for Meeting More Restrictive SO_2 Emission Limitations

Data collected by the Bureau of Mines during washability studies on 455 U.S. coal samples indicate that an additional 10 percent of these samples could meet an emission level of 0.52 $g/10^6$ J (1.2 lb $SO_2/10^6$ Btu) if they were physically cleaned by methods that allow recovery of 90 percent of the coal's heat value.[2,3] Insufficient samples were taken to show the variability in coal sulfur content, or to determine average sulfur contents over a given period of time. More extreme cleaning methods (such as finer grinding) would yield lower sulfur contents, but at the cost of lower heating value. Figure 2-3, which summarizes the Bureau of Mines study, indicates that the promulgation of a lower SO_2 emission standard would decrease to a small

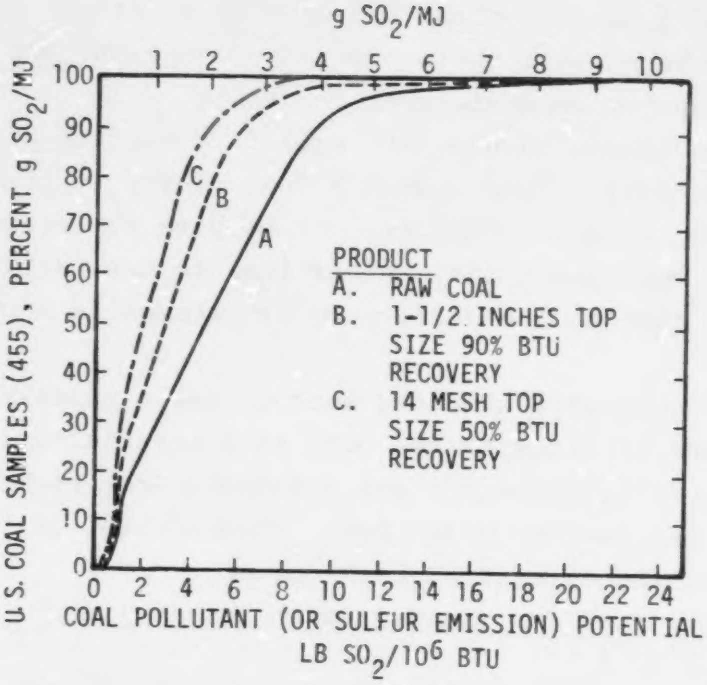

Figure 2-3. Effect of cleaning variable on coal SO_2 emission potential.[2]

percentage the amount of U.S. coal that can be burned, even after cleaning, without some additional form of emission control.

Another important fact determined from these studies is that the final sulfur levels to which coals can be cleaned vary from one coal region to another, from one coal bed to another within the same region, and, to a lesser extent, from one location to another within the same mine. Generalities are, however, difficult to make, and each coal seam should be assessed based on data for that area. These differences in cleaning potential result from variations in organic and pyritic sulfur levels and in the morphology of the coal-pyrite matrix, thus making it difficult to achieve a specified short-term average sulfur level in coal.

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2. Kilgroe, J.D. Implementation of Coal Cleaning for SO_2 Emission Control. In: EPA Program Conference Report - Fuel Cleaning Program: Coal. EPA-600/7-76-024. October 1976. pg. 57-65.

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2.3 EMERGING TECHNOLOGIES

A number of systems are being developed for firing coal and reducing SO_2 emissions. Those deemed applicable to reduce SO_2 from fuel combustion are fluid bed combustion (FBC) systems; systems for generation of low- or medium-Btu fuel gas from coal; and systems for production of liquid fuel from coal. Table 2-2 shows the technologies and applications that have recently received the most attention. They have been the subject of several recent reviews.[1,2,3]

The high cost of synthetic fuels, possible environmental impacts from production and combustion of liquid fuels, and competition for synthetic fuels for other uses (as transportation and synthesis gas for the chemical industry, for example) will probably combine to limit severely the use of synthetic fuels in large boilers. At present, only FBC and coal gasification appear to have near-term (5-10 years) possibilities for commercial application to boilers. It is difficult, however, to rule out any of the energy systems in Table 2-2 with certainty. The ability of both fluid bed combustion and coal conversion systems to meet emission standards may determine their rate of development.

A number of studies are being conducted for EPA to summarize the status of these technologies and their ability to achieve decreased SO_2 emissions. Because of the complex nature of these emerging technologies, they are not discussed further in this report.

Table 2-2. PROCESSES FOR UTILIZING COAL TO REDUCE SO_2 EMISSIONS

Basic technologies	Application
Atmospheric pressure fluid bed boilers	Utility or industrial boilers
Pressurized fluid bed boilers	Combined gas cycles for generation of electricity
Gasification of coal (low and medium Btu gas)	Production of fuel for combined cycles for generation of electricity
	Production of fuel for boilers of conventional design
Liquefaction of coal	Production of fuel for combined gas cycles for generation of electricity
	Production of fuel for boilers of conventional design

REFERENCES FOR SECTION 2.3

1. Fennelly, Paul F., et al. Preliminary Environmental Assessment of Coal-Fired Combustion Systems. Environmental Protection Agency, Research Triangle Park, N.C. Contract No. EPA 600/7-77-054. May 1977. p. 9.

2. Iammartino, Nicholas R. Fluidized-Bed Combustion: A Better Way to Burn Coal? Chemical Engineering, June 1976. pp. 69-71.

3. Hirschfeld, F. What's Holding Up Coal Gasification? Mech. Eng. August 1977. pp. 32-37.

3.0 FLUE GAS DESULFURIZATION SYSTEMS

Flue gas desulfurization (FGD) is the process of removing sulfur oxides, primarily SO_2, from combustion gases. In this process, flue gases are contacted with an absorbent medium in either an absorber or a scrubber vessel. Typically, SO_2 reacts with the absorbent or dissolves in the solution to produce a slurry that contains dissolved or solidified sulfur compounds. A number of dry adsorption systems are also under active development.

FGD processes are grouped into two basic categories, regenerable or nonregenerable, based on whether the sulfur compounds are separated from the absorbent as a by-product, or discarded along with the absorbent as waste. Nonregenerable processes produce a sludge that requires disposal in an environmentally sound manner. Regenerable processes include additional steps to convert the sulfur into by-products such as liquid sulfur dioxide, sulfuric acid, or elemental sulfur.

Six basic types of FGD systems are used in this country on large, coal-fired boilers. Of these, the lime scrubbing, limestone scrubbing, sodium carbonate scrubbing, and dual alkali systems come under the nonregenerable category. The magnesium oxide scrubbing system and Wellman-Lord system are regenerable. The following sections deal with these systems. Specific installations are described in detail, as well as their efficiency and operability, the problems involved, and their solutions.

3.1 LIME SLURRY FLUE GAS DESULFURIZATION SYSTEMS

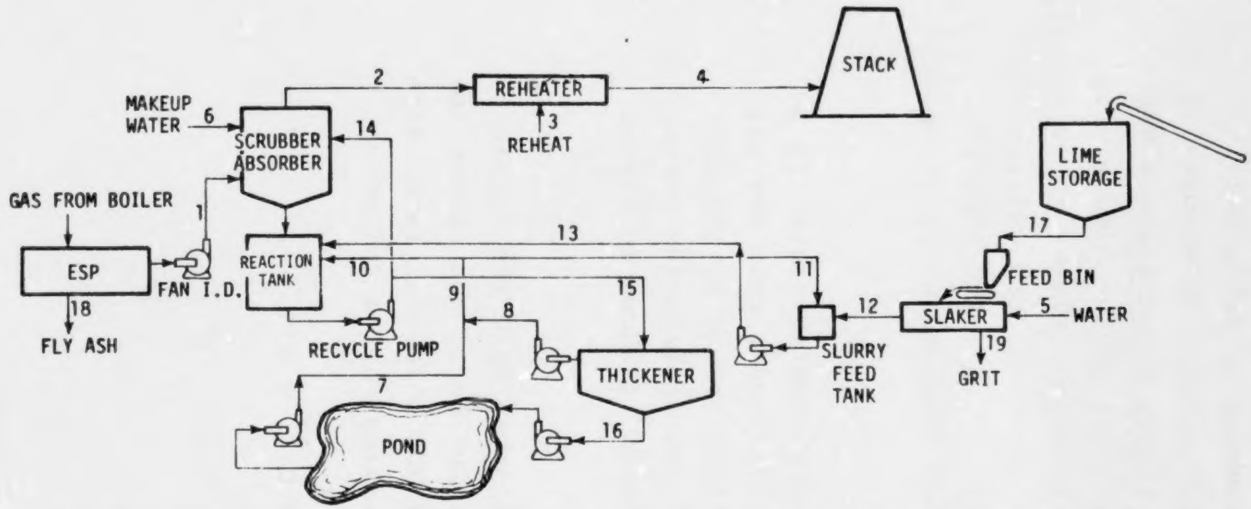
Lime slurry scrubbing is a wet scrubbing process that uses lime slurry to remove SO_2 from flue gas. Dry calcium oxide (CaO) is fed to the system and reacted with water to form a slurry. This slurry is contacted with the flue gas and absorbs SO_2. The SO_2 reacts with the slurry to form calcium sulfite, which is removed from the system as a sludge.

3.1.1 Process Description

This section presents a process description for lime-based FGD scrubbing systems and is divided under four subheadings:

- Process Flow/Material Balance
- Process Chemistry
- Waste Disposal
- Water Balance

3.1.1.1 Process Flow/Material Balance - The flow diagrams for operating lime scrubbing FGD systems appear in the appendix of this report. A model plant shown in Figure 3-1 shows the flue gas flowing from the boiler into ESP's, where 99 percent of the particulate is removed. The flue gas then flows through an induced draft fan into the absorber module (stream 1). Lime is fed from a feed bin to a slaker, where it is slurried with fresh water, diluted with recycled water and fed into the reaction tank (stream 13). A bleed stream (15) from the recirculating slurry goes to a thickener, the underflow from which is pumped to the sludge pond. Pond water and overflow from the thickener are returned to the system (streams 7 and 8). Makeup water enters as mist eliminator wash and as fresh water feed to the slaker. In actual practice, fresh water would also be used at the pump seals, but it is not shown in this case. Clean gas from the



Description gas streams	1 Gas to scrubber	2 Flue Gas to reheat	3 Reheat	4 Gas to stack
Rate, 10^3 lb/hr	5570	5840	1380	7220
SCFM 10^3	1110	1110	299	1410
Temp., °F	320	125	400	175
Fly ash 10^3 lb/hr	0.5	0.5		0.5
SO_2, 10^3 lb/hr	25.5	2.55		2.55
H_2O, 10^3 lb/hr	201	490	12.4	502
N_2, 10^3 lb/hr	3970	3970	1050	5020
CO_2, 10^3 lb/hr	1030	1030		1030
O_2, 10^3 lb/hr	143	342	320	660

Description liquid streams	5 Water to slaker	6 Makeup water	7 Pond water return	8 Thickener overflow	9 Return water	10 Return water to scrubber	11 Return water to slurry tank	12 Slurry to feed tank	13 Slurry to scrubber	14 Scrubber recycle	15 Spent slurry to thickener	16 Spent slurry to pond
Rate, 10^3 lb/hr	69.6	285	20.2	300	320	258	62.4	92.8	155	39500	429	129
GPM	138	565	40.1	595	635	512	124	156	277	72000	782	199
Solids, %	0	0	0	0.5	0	0	0	25.0	15.0	12.0	12.0	40.0
Specific gravity	1.00	1.00	1.00	1.00	1.00	1.00	1.00	1.183	1.111	1.089	1.089	1.286
Temp., °F	60	60	60	125	120	120	120	110	120	125	125	125
$CaSO_3 \cdot 1/2 H_2O$ 10^3 lb/hr										3340	41.7	41.7
$CaSO_4 \cdot 2H_2O$ 10^3 lb/hr										568	6.17	6.17
H_2O 10^3 lb/hr	69.6	285	20.2	300	320	258	62.4	69.6	112	14800	378	77.5

Description solid streams	17 Lime to slaker	18 Fly ash	19 Grit
Rate, 10^3 lb/hr	23.4	45.5	0.234

Figure 3-1. Flow diagram/material balance - lime slurry scrubbing systems for a 500-MW boiler.

scrubber absorber is reheated, using heated air, and discharged to the stack (stream 4). The assumptions needed to make this material balance are presented in Table 3-1. The material balance for this system appears in Figure 3-1.

3.1.1.2 Process Chemistry - Lime Slurry Preparation - The first step of the process is preparation of the lime slurry. Lime (CaO) is reacted with fresh water in a slaker to produce a slurry of calcium hydroxide ($Ca(OH)_2$) and water, according to the following reaction:[1] $CaO + H_2O \rightarrow Ca(OH)_2$ + heat. This material is diluted with recycled water to about 15- to 20-percent solids.

SO_2 Absorption Mechanism

Droplets of slurry pass countercurrently (usually) to the flue gas containing SO_2. The SO_2 passes from the gas into the droplets through an interface driven by the concentration gradient. The vapor-liquid equilibrium curve for SO_2 and a 10-percent lime solution is shown in Figure 3-2.

SO_2 Reactions

The SO_2 molecules that are absorbed into the liquid droplet must be removed from the solution if the absorption process is to continue. Dissolved SO_2 molecules react with the calcium species and water according to the following reactions:[2]

$$Ca(OH)_{2(s)} \leftrightarrow Ca(OH)_2\ (aq) \leftrightarrow Ca^{++} + 2OH^-$$

$$2OH^- + CO_2 \leftrightarrow CO_3^{--} + H_2O$$

$$CO_3^{--} + CO_2 + H_2O \leftrightarrow 2HCO_3^{--}$$

$$Ca^{++} + CO_3^{--} \leftrightarrow CaCO_3$$

$$CaCO_{3(s)} \leftrightarrow CaCO_3\ (aq) \leftrightarrow Ca^{++} + CO_3^{--}$$

$$SO_2\ (g) \leftrightarrow SO_2\ (aq) + H_2O \leftrightarrow H^+\ HSO_3^-$$

$$HSO_3^- \leftrightarrow H^+ + SO_3^{--}$$

Table 3-1. ASSUMPTIONS USED IN THE MATERIAL BALANCE CALCULATIONS FOR A 500-MW BOILER

SO_2 removal in absorber	90 percent
Coal sulfur content	3.5 percent
Coal heat value	27,900 kJ/kg (12,000 Btu/lb)
Lime CaO content	95 percent
Stoichiometry	1.1 percent
Sludge oxidation ($CaSO_3$/$CaSO_4$)	90/10
Fly ash/bottom ash	80/20
Particulate removal in ESP	99 percent
Excess air to air heater	40 percent
Excess air to boiler	10 percent
L/G	5.3 l/m^3 (40 gal/1000 scf)

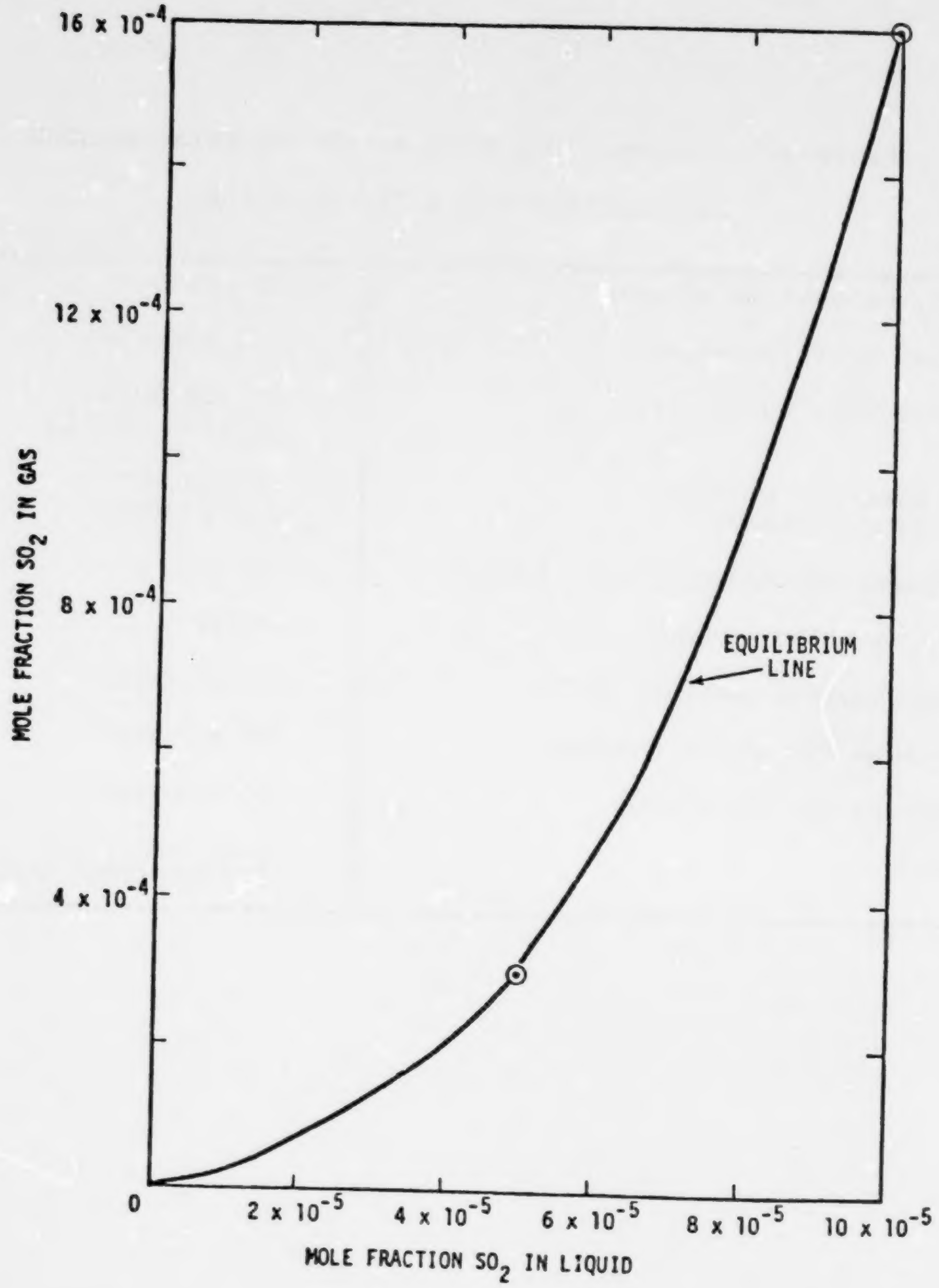


Figure 3-2. Vapor liquid equilibrium diagram SO_2 - 10 percent CaO slurry.

$Ca^{++} + SO_3^{--} \leftrightarrow CaSO_3(aq)$

$Ca^{++} + SO_3^{--} + 1/2\ H_2O \leftrightarrow CaSO_3 \cdot 1/2\ H_2O_{(s)}$

$CaSO_3 + H^+ \leftrightarrow Ca^{++} + HSO_3^-$

$H^+ + HCO_3^- \leftrightarrow H_2CO_3 \leftrightarrow CO_{2(g)} + H_2O$

$CaSO_3 \cdot 1/2\ H_2O + 3/2\ H_2O + 1/2\ O_2 \leftrightarrow CaSO_4 \cdot 2H_2O$

All the above reactions take place to some extent in every lime/limestone scrubbing system; they depend on many factors, including ionic concentrations, pH, temperature, and retention time in the reaction tank. The net effect of the reactions is the removal of SO_2 from the flue gas, depletion of calcium hydroxide in the slurry, and the generation of calcium sulfite in the slurry. The calcium ions used in the reactions are replenished by dissociation of the calcium hydroxide.[3] This is a primary difference between the lime and limestone slurry scrubbing systems and the so-called clear solution scrubbers (such as double alkali scrubbers), where the predominant reactive alkali (sodium) is more soluble.

In order to complete the SO_2 removal reactions listed above, a certain amount of retention time is needed for the slurry to achieve chemical equilibrium before it is recirculated to the scrubber. To do this, a reaction tank is usually used with a retention time of between 5 and 20 minutes.

3.1.1.3 Waste Disposal - The calcium sulfite and calcium sulfate that are formed in the absorber and the absorber reaction tank must be removed from the system. This is usually accomplished by bleeding a stream from the recirculating slurry to a thickener or clarifier. The calcium sulfite and sulfate settle out and are removed from the underflow of the clarifier in a slurry of about 30-percent solids.[4] The overflow, a clear liquid of less than 1-percent solids, is brought back to the system.[5] The sludge from the underflow is disposed of by several means.

One method of disposal is to pump the sludge to a pond. In this case the sludge is pumped from the thickener underflow (or occasionally directly from the reaction tank) to a large settling pond, where the solids settle. Clear water is recycled from the pond, as necessary, to keep the pond from overflowing. Return of all this clear liquor from the pond is called a "closed-loop" operation.

Another method of disposal is to dewater the sludge in a vacuum filter, chemically fixate it and dispose of it in a landfill. In this case there are commercially available processes to stabilize the sludge, such as the IUCS system.[6] When vacuum filtration is used, the filtrate is returned to the system to create closed-loop operation.

3.1.1.4 Water Balance - To minimize liquid discharge, closed-loop operation is desirable, and maintaining a water balance in the scrubbing system is very important. The quantity of water added can not exceed the amount lost in the flue gas and in the sludge. The moisture content of the flue gas is increased by cooling the incoming flue gases from about 149°C (300°F) to their adiabatic saturation temperature of about 53°C (128°F); this is achieved by evaporating the water in the scrubbing system.[7] Water is lost in the sludge both as water of hydration and as free water in the sludge. Water is added to the system as slaker feed water, pump seal water, and mist eliminator wash water.

3.1.2 Domestic Lime Scrubbing Units

The first major domestic lime FGD system, the Phillips Power Station of Duquesne Light Co., started up in 1972.

Since that time, 11 other major stations have installed lime scrubber systems (Table 3-2). A description of each of these is given in the Appendix A.

It is important to note that each of these facilities that has been tested is operating at or above the efficiency required to meet SO_2 emission legislation.[8]

A more thorough review of three units, Green River, Bruce Mansfield, and Mohave, is presented in the following pages. Respectively, these units represent high FGD system availability, a large system with high SO_2 removal efficiency, and a unit with very high SO_2 removal efficiencies on a low-sulfur coal application.

3.1.2.1 Green River Facility Description - The Green River Station of Kentucky Utilities is on the Green River in central Kentucky, approximately five miles north of Central City. The plant contains four steam turbine generating units having a total gross generating capacity of 242 MW. Boilers 1, 2, and 3, the boilers with the FGD units, supply steam for two steam turbine generators with a combined generating capacity of 64 MW. These two electrical generating units are used for peak loads and normally operate on a 5-day week, with one or more of the boilers often at reduced capacity.

All three boilers are dry-bottomed and burn pulverized coal. They were manufactured by Babcock and Wilcox and put into service in 1949 and 1950. There are no plans to retire them.[10]

The utility purchases a high-sulfur coal, which is used in conjunction with the FGD system. This coal comes from the Drake Mine in Muhlenberg County, Kentucky, and is shipped to the plant via barge. A typical analysis of it gives the following values: heating value, 25,100 kJ/kg

Table 3-2. LIME BASED FGD SYSTEMS IN THE UNITED STATES[9]

Station	Company	Size, (MW)	Start-up date	Status	New or retrofit	Coal, % sulfur	Design SO_2 removal efficiency, %
Bruce Mansfield	Pennsylvania Power Co.	835	April 1976	Operational	N	4.7	92
Cane Run	Louisville Gas & Electric	178	August 1976	Operational	R	3.5-4	85
Colstrip	Montana Power Co.	360	December 1975	Operational	N	0.8	60
Conesville	Columbus & Southern Ohio Electric Co.	400	February 1977	Operational	N	4.5-4.9	90
Elrama	Duquesne Light Co.	510	October 1975	Operational	R	1-2.8	85
Four Corners	Arizona Public Service	160	February 1976	Test facility shutdown	R	0.7	N.A.
Green River	Kentucky Utilities	64	September 1975	Operational	R	3.8-4	80
Mohave	So. California Edison	160	November 1973	Test facility shutdown	R	0.8	N.A.
Paddy's Run	Louisville Gas & Electric	65	April 1973	Operational	R	3.5-4	80
Phillips	Duquesne Light Co.	410	April 1972	Operational	R	1-2.8	83
Rickenbacker	USAF	20	March 1976	Operational	R	3.6	90
Shawnee	TVA/EPA	10	April 1972	Test facility	R	0.8-5.0	N.A.

(10,800 Btu/lb); sulfur content, 3.7 percent; ash content, 13.4 percent; total moisture, 12.1 percent.[11]

Boilers 1, 2, and 3 are fitted with mechanical collectors upstream of the FGD system. Design particulate removal efficiency is 85 percent. The FGD system was designed and installed by American Air Filter (AAF) and consists of one scrubber module to handle a maximum flue gas capacity of 170 m^3/s (360,000 acfm) at 149°C (300°F). Table 3-3 presents pertinent plant design and operating data.[12,13]

The process is conveniently described in terms of two basic operations: the flue gas scrubbing system, and the lime slurry/recycle system. Figure 3-3 is a schematic flow diagram of the process.

Flue Gas Scrubbing System

The flue gas from each boiler coupled into the scrubbing system initally passes through a series of mechanical collectors (Western Precipitation, multicyclone, 9-in. diameter, cast iron construction) where primary particulate removal takes place. The flue gas is then drawn from the existing breeching through a guillotine-type isolation damper and associated ductwork to the scrubber fan. The incorporation of the guillotine-type isolation dampers allows selective bypassing of the flue gas around the scrubbing system to an existing stack.[14]

Before entering the scrubbing system, the flue gas passes through a booster fan with a power rating of 1118 kW (1500 Hp) and a pressure increase of 0.46 m (18 in.) H_2O. From the outlet of the scrubber booster fan, the gas flows through a variable-throat flooded elbow venturi scrubber. The venturi scrubber was installed to control particulate emissions from the mechanical collectors and to quench the hot gas. Quenching lowers the temperature of the inlet gas

Table 3-3. POWER PLANT AND FGD SYSTEM DESIGN DATA[16]

Green River - Kentucky Utilities

Boiler data	Generating capacity, MW	64
	Year placed in service	1949
	Boiler manufacturer	Babcock & Wilcox
Coal data	Heat value	25,100 kJ/kg (10,797 Btu/lb)
	Ash content	13 to 14 percent
	Sulfur content	3.8 percent
FGD system data	SO_2 removal efficiency	80 percent design
	Particulate removal efficiency	99.7 percent
	Start-up date	9/75
	Flue gas rate	170 m^3/s (360,000 acfm)
	Flue gas temperature	149°C (300°F)
	Stack height	24 m (78 ft)
	FGD vendor	American Air Filter

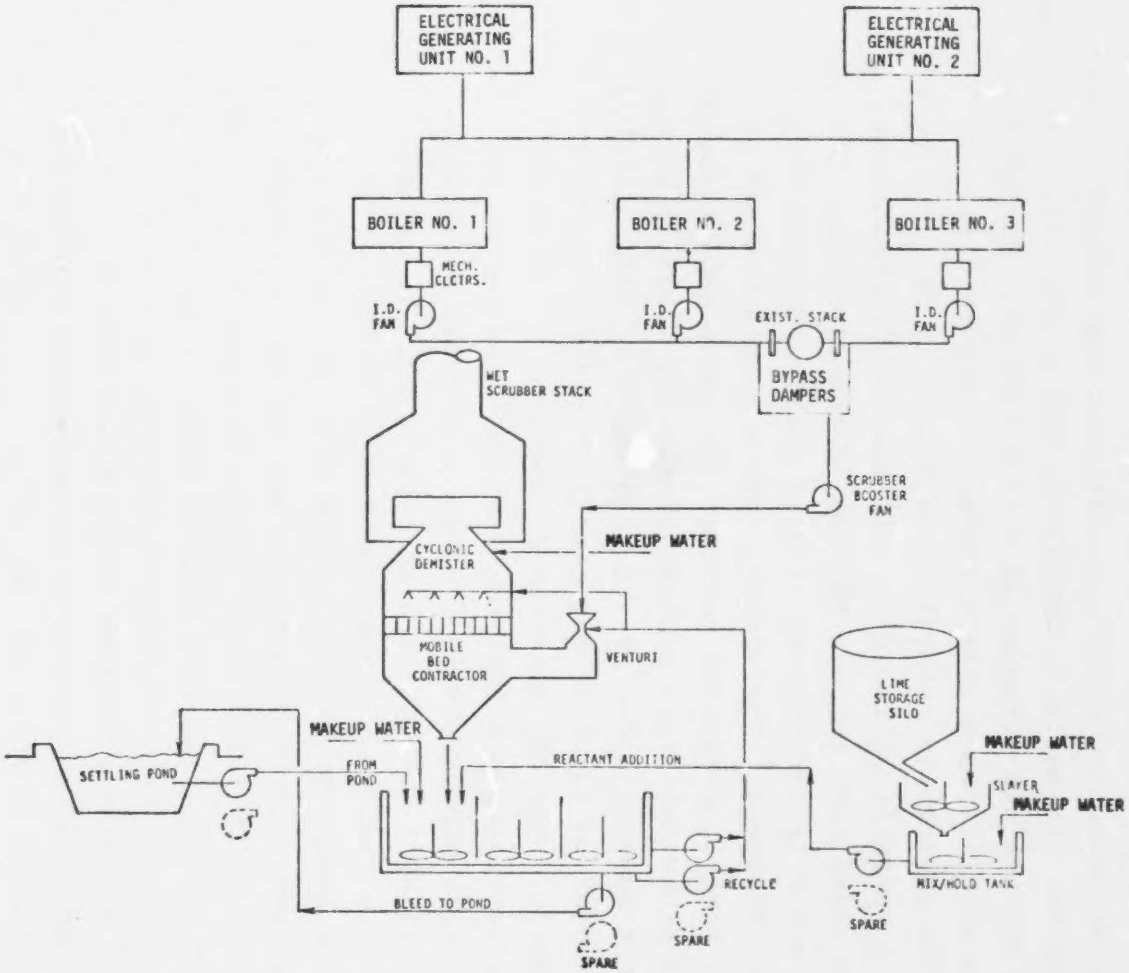


Figure 3-3. Process flow diagram - Green River No. 1, 2, and 3.[15]

from 163°C (325°F) to approximately 47°C (116°F) within the scrubber module. This substantially decreases the volume of gas to be scrubbed and provides additional temperature protection for the balls in the mobile bed contactor.

The venturi has a throat of 2.54 m (100 in.) diameter operating with a 2.29-m (94-in.) plug. Pressure drop through the venturi is maintained at 0.18 m (7 in.) H_2O by a Limitorque operator on the plug. Liquid flow is maintained at 83 l/s (1360 gpm) through the top of the scrubber. The scrubber shell is constructed of mild steel and lined with acid-proof Precrete. The venturi throat is constructed of stainless steel. After the venturi, the gas passes through a flooded elbow and flows upward through the mobile bed contactor at a rate of 136 m^3/s (288,200 acfm) at 47°C (116°F). Sulfur dioxide in the flue gas reacts with the lime slurry at this stage. The SO_2 absorber consists of a mobile bed contactor section and a demister section. The mobile bed stage contains approximately 175,000 to 190,000 solid polyethylene balls of 3.2 cm (1.25 in.) diameter. It is constrained to a maximum thickness of 0.61 m (2 ft), or 0.41 m (16 in.) at rest. The slurry is fed at a rate of 595 l/s (9750 gpm) and applied both to the bed and to the upward rising flue gas by 125 five-cm (2-in.) overhead nozzles and sixty-five 3.2-cm (1.25-in.) nozzles spraying upward into the bed. The contactor bed is compartmentalized into 10 individual sections. Underbed dampers are used to meet flue gas turndown requirements. Pressure drop through the contactor bed is 10 cm (4 in.) H_2O.[17]

Following passage through the bed, the gases continue upward to a single stage, spin vane demister. The demister is continuously washed by spray nozzles at a rate of 3.1 l/s

(50 gpm). Pressure drop is 2 in. H_2O.[18]

The scrubbed flue gas (140 m^3/s @ 47°C or 296,300 acfm @ 116°F) is then discharged to the atmosphere through a wet stack. This stack is constructed of carbon steel and lined with Carboline.

Lime Slurry/Recycle System

The scrubbing slurry feed and recycle system consists of a partitioned concrete reactant tank, recycle pumps to supply the scrubber and absorber module, a lime slurry slaking and feed system, a bleed system for scrubbing waste discharge to a settling pond, and a return water system for recycling the water from the settling pond back to the process.

The reactant tank, constructed of acid-proof concrete, is 22 m (72 ft) long; 7.3 m (24 ft) wide; and 7.3 m (24 ft) high. Two partitions form three individual compartments, each agitated and connected by underflow openings. The total liquid capacity of each compartment is 378.5 m^3 (100,000 gal.). The total retention time in the reactant tank system is 21 minutes--7 minutes per compartment.[19]

From the reactant tanks, recycle pumps feed both the venturi particulate scrubber and the mobile bed contactor. These pumps (two operational, one spare) are rated at 360 l/s (5900 gpm) each. All pumps and agitators are rubber-lined.[20]

Reaction products and collected particulate matter are pumped to an impervious clay-lined pond on the plant site approximately 805 m (1/2 mile) from the scrubbing module. The pond has a capacity of 182,630 m^3 (148 acre ft) and a depth of 6.1 m (20 ft). River water is used as makeup and is introduced into the reactant tank, lime slaking tank, and demister, as well as the pump seals. Total fresh water makeup supplied to the system is 4.6 l/s (75 gpm).[21,22]

Operating History

Commercial operation of the scrubbing system commenced in the fall of 1975. Before commercial service, the system was put through an extensive four-phase prestart-up testing program, which included mechanical and electrical trials operation on air and water, verification of mechanical reliability, and hot flue gas operation.

The operation on flue gas began September 13, 1975. It was conducted on a 50-percent load basis with 1.6- to 2.0-percent sulfur coal. During this phase, several minor problems were uncovered and corrected. The major areas of concern included the pH sensors and SO_2 analyzers. In addition, some problems were encountered with plugged spray nozzles.

Closed-loop, full capacity operation began in March 1976 with the initiation of a 6-month, vendor qualification operation period.[23] To date, the performance of the system has been good. Mechanical reliability has been excellent. The index values for average system availability, operability, reliability, and utilization (including start-up) for 1976 are presented in Table 3-4. The operability is plotted through May 1977 (Figure 3-4).

Based on continuous monitoring data recorded by AAF, SO_2 removal efficiency has been well above the design guarantee value, at about 90 percent.[24]

Problems and Solutions

Some chemistry-related problems have been encountered. Specifically, a hard coat of scale developed in the lower sections of the absorber internals. This was believed to have been caused by a high scrubbing solution pH combined with high oxygen concentrations in the flue gas. Chemically, this happened as follows: As the pH of the scrubbing solu-

Table 3-4. GREEN RIVER POWER STATION OPERATIONAL DATA FGD UNIT[25,26] FOR 1976 AND 1977 (SEPTEMBER)

		Parameters								
Year	Month	A Hours in period	B Hours FGD system available	C Hours FGD called upon	D Hours FGD system operated	E Hours boilers operated	B/A % avail-ability	D/C % Reli-ability	D/E % oper-ability	D/A % utiliza-tion
1976	January	744	312.00	456.00	64.00	571.55	41.9	14.0	11.2	8.6
	February	696	486.17	499.38	210.75	499.38	69.9	42.2	42.2	30.3
	March	744	721.72	408.66	386.38	457.53	97.0	94.5	84.4	51.9
	April	720	648.00	552.00	552.00	552.00	90.0	100.0	100.0	76.7
	May	744	606.18	455.88	455.88	455.88	81.4	100.0	100.0	61.2
	June	720	720.00	596.43	588.85	596.43	100.0	98.7	98.7	81.8
	July	744	665.85	583.53	574.43	583.53	89.5	98.4	98.4	77.2
	August	744	722.45	744.00	722.45	744.00	97.1	97.1	97.1	97.1
	September	720	617.20	571.20	571.20	571.20	85.7	100.0	100.0	79.3
	October	744	744.00	698.55	698.55	698.55	100.0	100.0	100.0	93.9
	November	720	720.00	704.25	704.25	704.25	100.0	100.0	100.0	97.8
	December	744	539.31	591.48	517.20	535.52	72.5	87.4	96.6	69.5
		8784	7502.88	6861.36	6045.94	6969.82	85.4	88.1	86.7	68.8
1977	January	744	698.29	744.00	698.26	744.00	93.9	93.9	93.9	93.9
	February	672	242.80	266.12	242.80	266.12	36.1	91.2	91.2	36.1
	March	744	0	0	0	0	0	0	0	0
	April	720	288.00	166.82	164.00	166.82	40.0	98.3	98.2	22.8
	May	744	735.65	526.55	513.27	526.55	98.9	97.5	97.5	69.0
	June	720	720.00	34.38	34.38	34.38	100.0	100.0	100.0	4.8
	July	744	744.00	0	0	0	100.0	0	0	0
	August	744	744.00	0	0	0	100.0	0	0	0
	September	740	740.00	0	0	0	100.0	0	0	0

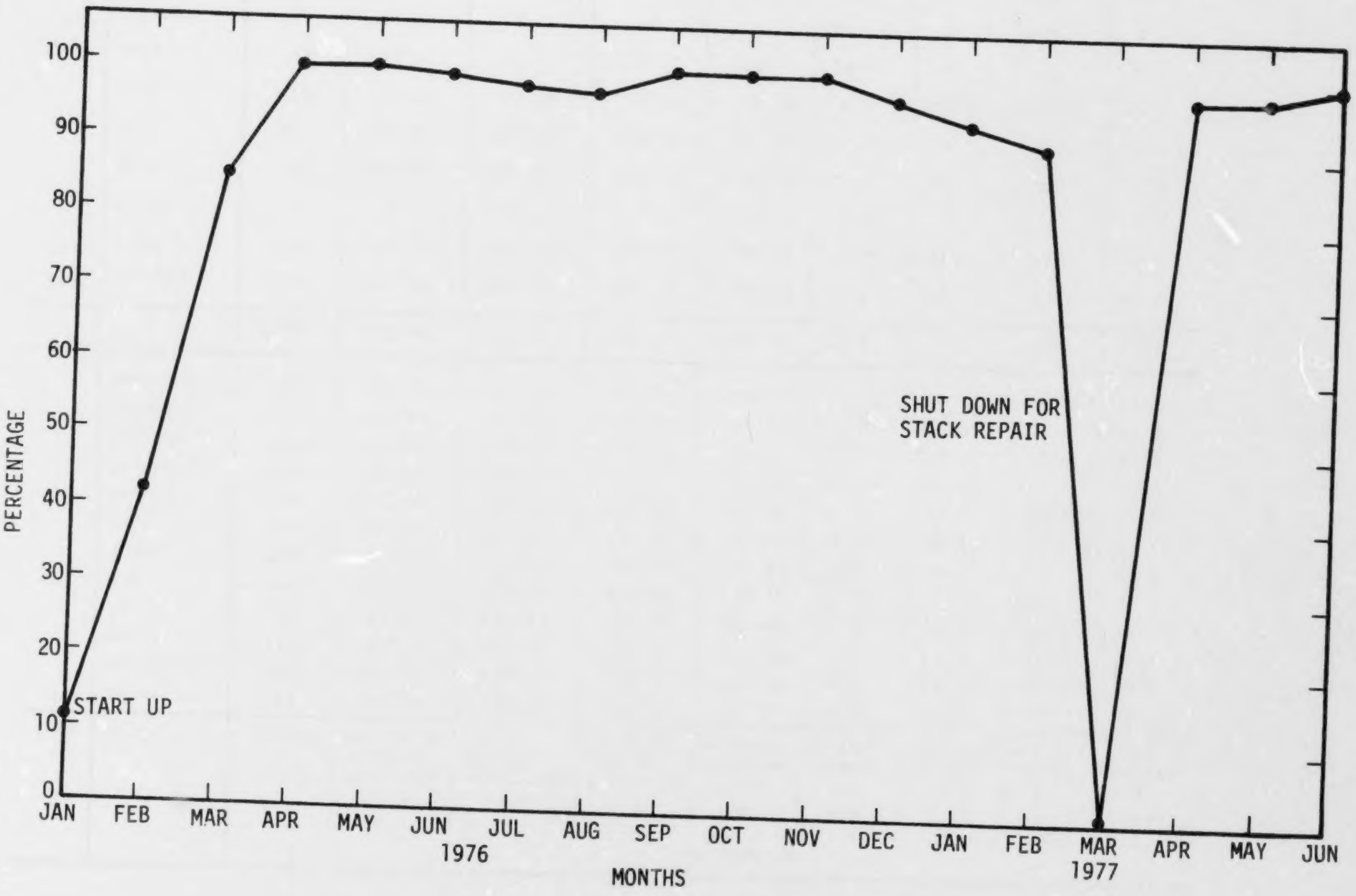


Figure 3-4. Scrubber system operability - Green River No. 1, 2 and 3.[27]

tion became more alkaline, in the range from 9 to 10, calcium sulfite tended to precipitate. In the presence of high oxygen concentration, the sulfite was chemically oxidized to sulfate, resulting in the formation of a hard $CaSO_4$ scale. This problem was corrected by manually removing the scale, lowering the oxygen content of the flue gas by minimizing air leakage, and relocating the pH sensors to a turbulent part of the absorber reaction tank. Following completion of these modifications, scale formation has not been a serious problem; and although some scale film has been periodically noted, this has tended to disappear with continued operation and periodic cleaning.[28]

Acid fall-out from the plume has resulted in the need to compensate a number of employees for damage to the finish on their automobiles. A pH as low as 2 has been found in moisture drained from the stack.[29]

The system was shut down for 65 days from February to April 1977, to cover one half of the inner stack wall for its entire height with 9.5-mm (3/8-in.) steel plate. The stack was then lined with 19-mm (3/4-in.) precrete over a wire mesh. The original lining of Carboline had failed around nearly half the circumference.

The utility is now replacing the radial vane demister with a chevron type to reduce the acid mist loading. To avoid further damage to automobiles and the possibility of damage to an adjacent substation, AAF has been authorized to design and install a stack gas heating system. The proposal is to use extraction steam from another unit; this with heat air to be injected into the stack and thus raise the stack temperature by some 10°C (50°F).[30,31]

3.1.2.2 Bruce Mansfield Facility Description - The Bruce Mansfield plant is a 2700-MW, three-unit, coal-fired facility on the Ohio River in the Borough of Shippingport, Pennsylvania. It was built by Pennsylvania Power Co., which is acting on its own behalf and as an agent for the other participating companies of the CAPCO consortium, namely, The Cleveland Electric Illuminating Co.; Duquesne Light Co.; Ohio Edison Co.; and Toledo Edison Co. All three boilers will burn 4.7-percent-sulfur coal and require FGD to meet emission standards. Unit No. 1 was started up in April 1976. Unit No. 2 is currently undergoing start-up and Unit No. 3 is scheduled for start-up in 1980. The remainder of this section discusses the Bruce Mansfield No. 1 facility.

Bruce Mansfield No. 1 is a coal-fired, once-through, supercritical steam generator which fires 302 Mg/hr (333 tons/hr) of coal and generates approximately 2.9 Mg/hr (6.5 10^6 lb/hr) of steam at 26.1 MPa (3785 psig) and 541°C (1005°F). Emission control equipment required for this unit is designed to meet state emission regulations of 25.85 $SO_2/10^8$ J (0.6 lb $SO_2/10^6$ Btu) and 0.04 g/m^3 (0.0175 gr/scf) of particulate, when burning 27,700 kJ/kg (11,900 Btu/lb) coal having average ash and sulfur contents of 12.5 and 4.7 percent respectively. Additional design related information is presented in Table 3-5.[32]

The emission control system is a venturi wet scrubber system for removal of sulfur dioxide and particulates. The system was designed and manufactured by Chemico Air Pollution Control Co. and uses the Dravo Corporation's thiosorbic lime as the scrubbing absorbent. A flow diagram is shown in Figure 3-5.

Each scrubbing train consists of a variable-throat venturi, a 6711-kW (9000-hp) I.D. fan, and a fixed-throat venturi absorber vessel. There are six scrubber trains, all of which are required to handle the flue gas flow from the

Table 3-5. POWER PLANT AND FGD SYSTEM DESIGN DATA[33]

Bruce Mansfield No. 1 - Pennsylvania Power Co.

Boiler data	Generating capacity	825 MW (net)
	Year placed in service	1976
	Boiler manufacturer	Foster-Wheeler Corp.
Coal data	Heat value	27,700 kJ/kg (11,900 Btu/lb)
	Ash content	12.5 percent
	Sulfur content	4.5 to 5.0 percent
FGD system	SO_2 removal efficiency	92 percent
	Particulate removal efficiency	99.8 percent
	Start-up date	4/76
	Flue gas rate	1580 m^3/s (3,350,000 acfm)
	Flue gas temperature	196°C (385°F)
	Stack height	290 m (950 ft)
	FGD vendor	Chemico

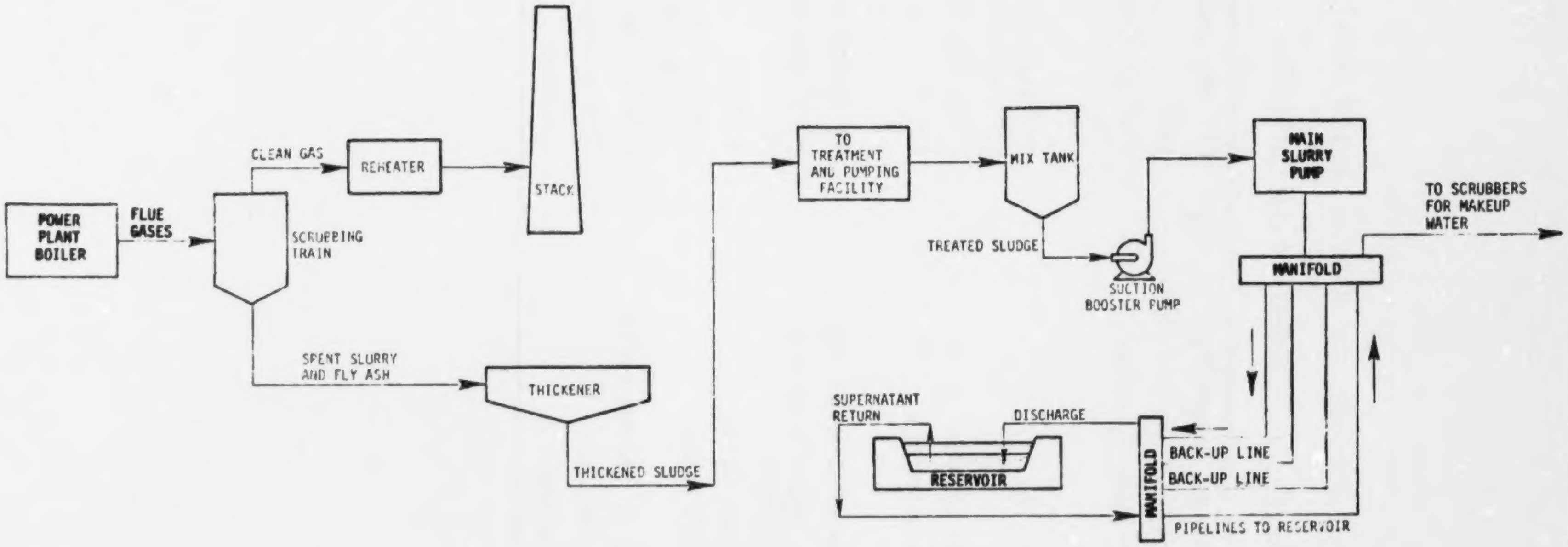


Figure 3-5. Simplified process diagram - Bruce Mansfield No. 1.[34]

boiler at full load. The scrubbers/absorbers are arranged in two groups of three trains. The scrubbed flue gas is ducted through a common reheat chamber (one for each three train group) and then exhausted through two separate flues in the 290-m (950-ft) stack.[35]

The adjustable-throat venturi scrubbing module removes nearly all of the fly ash contained in the flue gas; the absorber module removes most of the remaining fly ash. Sulfur dioxide is absorbed in both the scrubber and absorber by droplets of lime slurry, containing 2- to 6-percent magnesium oxide. The unit is guaranteed to remove 92 percent of the SO_2 and 99.8 percent of the particulate matter from the flue gas.[36]

The scrubber-recycle bleed is combined with a fly ash slurry from the boiler and discharged to a thickener of 61 m (200 ft) diameter. Sludge from the thickener underflow is pumped to a waste disposal system, where it is mixed with a stabilizing agent (Calcilox) and then pumped approximately 11 km (7 miles) to the Little Blue Run Ravine for landfill.[37]

The initial shakedown and debugging phase of operation began for part of the system in December 1975.[38] Full commercial operation began in April 1976.[39]

Performance History

This facility reported 98-percent operability during the first 7 months after commercial certification (June 1, 1976). Operating problems were solved without causing boiler downtime. However, during the exceptionally cold winter months of January and February 1977, the boiler lost 11 percent and 24 percent of generation respectively. On March 21, 1977 the system was taken down for a 10-week turbine overhaul. During this time, repairs also began on the chimney flue for A, B, and C modules. The polyester flakeglass lining had failed and was being replaced.

Previously a spray application had been used, but the new lining is being troweled on. The other chimney flue, for the D, E, and F modules, also needs repair. Roughly one year will be required to complete the work and the boiler will be held to half load for that time.[40] During the boiler and chimney outages, Pennsylvania Power will be making other system changes, as described in the following paragraphs.

Performance Test - Pennsylvania Department of Environmental Resources tested the unit July 19 and 20, 1977. Sulfur dioxide emissions were 19 g/10^8 J (0.44 lb $SO_2/10^6$ Btu) and 54 g/10^8 J (1.26 lb/10^6 Btu), respectively. This represents 89- and 78-percent removal. The allowable emission rate is 26 g/10^8 J (0.6 lb/10^6 Btu). The details and causes for such a wide variation in results have not been published.[41]

Operating Problems and Solutions

Mist Eliminator - The major problem is with the four-pass, chevron-type, horizontal mist eliminator. The company has experienced mist carry-over during scrubber operation. The mist eliminator is designed for liquid carry-over of 2.3 g/m^3 (1 gr/scf), but plant personnel estimate an actual carry-over of about 6.9 g/m^3 (3 gr/scf). The eliminator was designed to operate at gas velocities of 2.4 to 3 m/s (8 to 10 ft/s). Pennsylvania Power would like to install a new vertical mist eliminator in the ducting downstream of the scrubber vessels; because of duct diameter and space restrictions, however, such a design would necessitate flow velocities as high as 15.2 m/s (50 ft/s). A plastic vertical mist eliminator was installed, but collapsed as a result of structural failure. The problem of plugging was also apparent, but has been solved by relocating the lime feed to the scrubber at a lower position.[42,43]

Closed-Loop - The system is not closed-loop, since water is being retained at the sludge disposal site and not recycled to the process. Makeup water from the disposal pond is not needed, because "fresh" water is added to the system as fan sprays and for lime slaking. The company thinks it can operate closed-loop; it has not done so because of its preoccupation with the system's mechanical problems. In the event it does not operate closed-loop, however, the plant has a permit to discharge water to the Ohio river. Discharge requirements for water are: total suspended solids of 60 mg/l or less; aluminum content of 10 mg/l or less; and a pH between 6 and 9.[44,45]

I.D. Fans - Pieces of scale were carried through the system; the rubber lining on the fan scrolls was chipped away, resulting in exposure and corrosion of the carbon steel. These fan scrolls are now being replaced with ones made of Inconel. Pitting has also occurred on the carbon steel hubs, the result of seal leakage. The fan shafts are made of carbon steel clad with Carpenter 20 alloy.[46]

Reheater - The reheaters are not used because of duct vibrations caused by resonant conditions. They are now undergoing design changes to correct the problem. Without reheat, plume buoyancy appears sufficient, but atmospheric conditions can occasionally cause condensation and precipitation of moisture from the plume. Winds are such that the liquid fallout occurs over the nearby town of Shippingport. This fallout is a clear liquid, but on drying leaves a film on automobiles, windows, and other surface. The pH of the condensed liquid is about neutral.[47]

Scrubber - Some plugging of the venturi nozzles with scale has occurred. This problem was eliminated by installation of "baskets" in the main header to catch the scale particles upstream of the nozzles. The baskets can be cleaned without shutting down the system. Maintenance of the scrubber system is not performed on a routine basis, but only when needed. Scheduled maintenance occurs every 12 months.[48]

pH - pH had to be manually sampled and controlled because of difficulties with the automatic system. This system is being redesigned by plant personnel.[49,50]

3.1.2.3 Mohave Generating Station Facility Description - As part of their compliance program, the participants of the Navajo/Mohave Power Project funded a prototype scrubber demonstration project at the Mohave Generating Station. Participants in the project are:[51]

Salt River Project Agricultural Improvement and Power District
Arizona Public Service Company
Department of Water and Power of the City of Los Angeles
Nevada Power Company
Tucson Gas and Electric Company
Bureau of Reclamation of the United States Department of the Interior
Southern California Edison Company

Design of the 170-MW demonstration facility was based on the results of an extensive pilot plant program in which different scrubber configurations were tested, (Table 3-6). Two types of absorber system were built for the demonstration tests. One was a horizontal cross-flow scrubber (Figure 3-6) and the other a vertical counter-current flow system. The horizontal module was designed to treat 212 m^3/s (450,000 scfm) of flue gas. In cross-section the unit was 4.6 m (15 ft) high by 8.5 m (28 ft) wide. It was 14.6 m (48 ft) long. A 1305-kW (1,750-hp) booster fan supplied flue gas to the scrubber from the discharge of the ESP.

Table 3-6. POWER PLANT AND FGD SYSTEM DESIGN DATA[53]

Mohave Test Plant - Southern California Edison

Boiler data	Generating capacity	790 MW
	Year placed in service	1971
	Boiler manufacturer	Combustion Engineering
Coal data	Heat value	26,800 kJ/kg (11,500 Btu/lb)
	Ash content	10 percent
	Sulfur content	0.8 percent
FGD system	SO_2 removal efficiency	95 percent
	Particulate removal efficiency	93 percent
	Start-up date	11/73
	Flue gas rate	212 m^3/s (450,000 scfm)
	Flue gas temperature	149°C (300°F)
	Stack height	152 m (500 ft)
	FGD vendor	Southern California Edison Stearns Roger

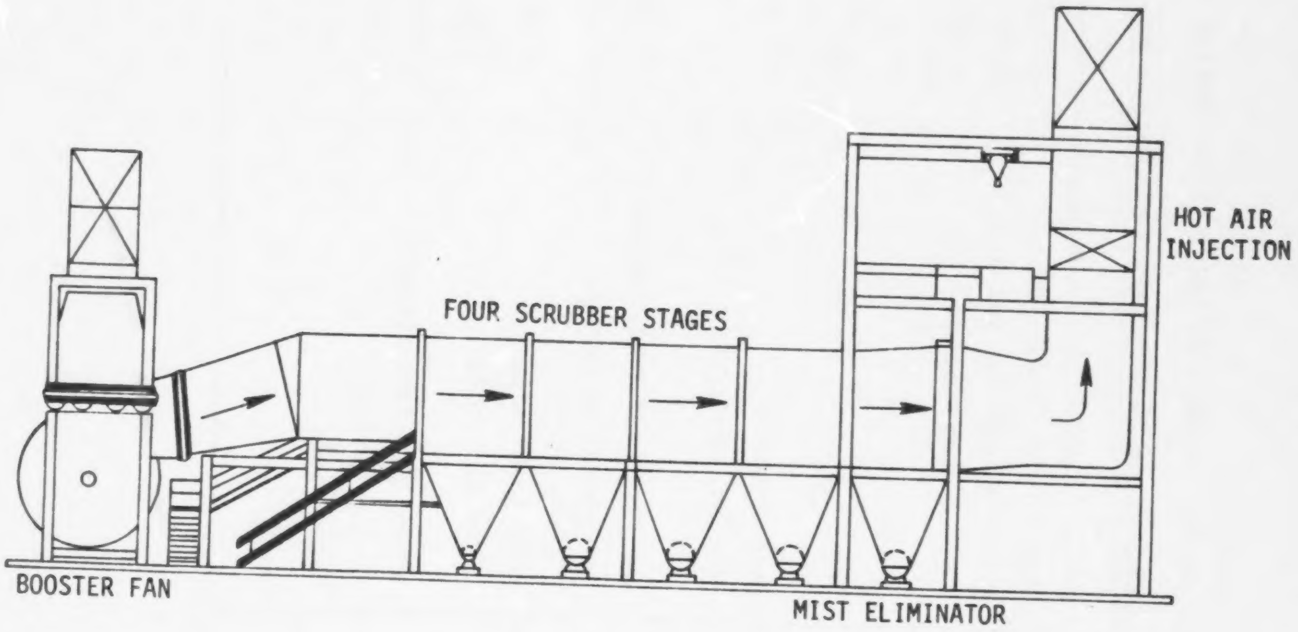


Figure 3-6. Horizontal test module Mohave plant.[54]

The demisters were in the discharge side of the absorber and reheat was provided by externally heated ambient air. The flue gas was discharged through a 152-m (500-ft) stack. The module had four stages with two slurry pumps and one spare for each stage. Flow was 549 l/s (9000 gpm) through 36 nozzles per stage. The slurry was sprayed perpendicularly to the gas flow (i.e. cross-current) and was recycled from stage to stage countercurrent to gas flow.[52]

This horizontal module was operated from January 16, 1974, until February 9, 1975, and was tested more extensively than the vertical module. The dewatering system consisted of a thickener tank and a sludge disposal pond. Water from the pond was returned to the system.

The vertical module, which consisted of either a TCA or packed grid (PPA), was also rated at an equivalent of 170 MW, but this system was operated mainly on limestone (see Section 3.2).[55]

Performance Results - Sulfur dioxide removal efficiency was excellent for all three types of absorbers. Although the SO_2 inlet concentration was only 200 ppm, all three types of absorber were capable of removing 95 percent of the inlet SO_2. Outlet SO_2 loadings ranged from 1 to 10 ppm range.[56]

The SO_2 removal efficiency was strongly dependent on L/G for all three modules described in Section 3.1.4 and shown in Figure 3-7. Note that the L/G shown for the horizontal module represents the ratio in each stage.

Sulfur dioxide removal efficiency was directly related to the number of contact stages employed. These data are shown in Section 3.1.4.5 for the horizontal module. The first stage removed 50 percent of the SO_2. Two-stage removal

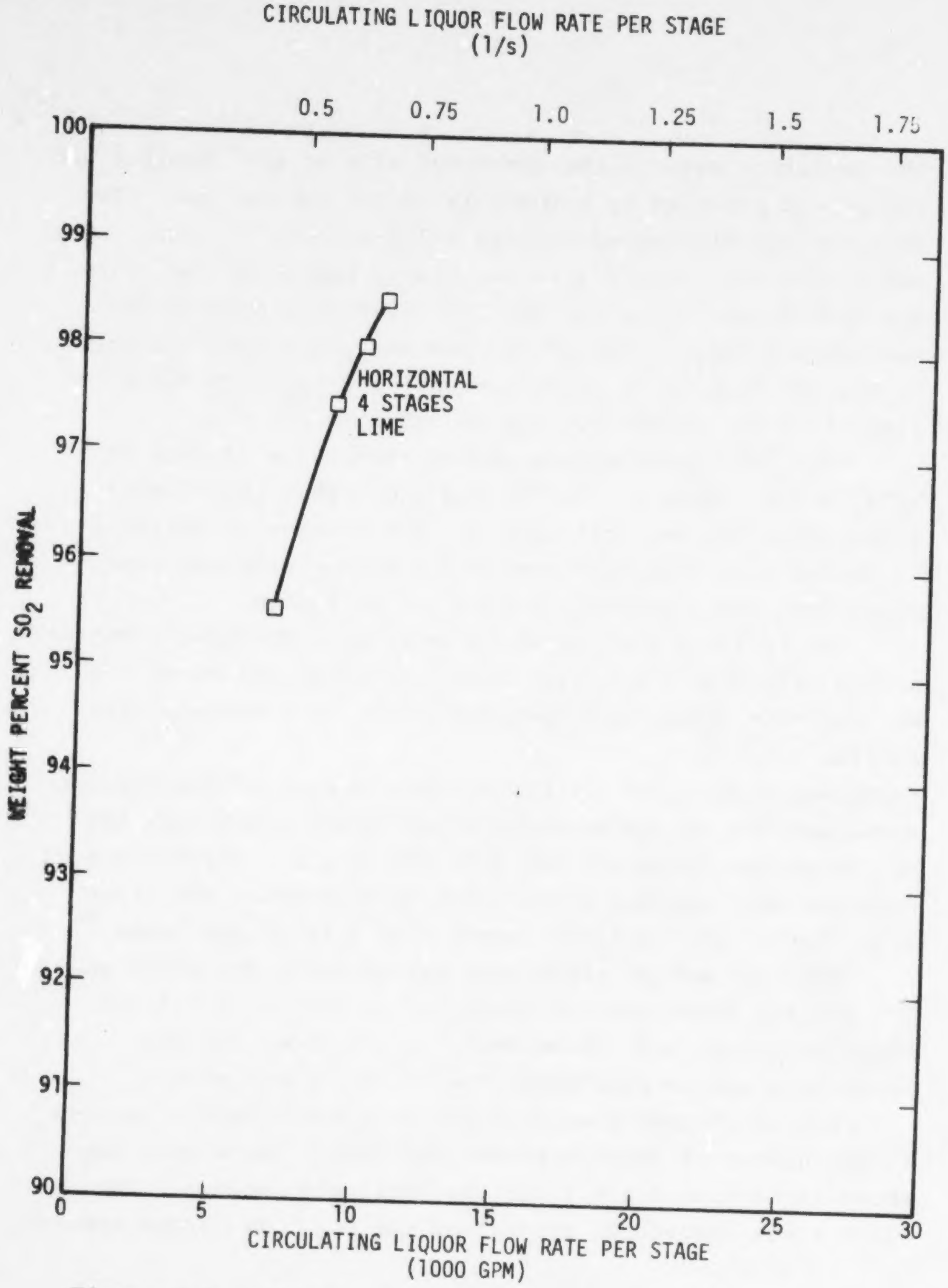


Figure 3-7. Effect of circulating liquor flow rate on SO_2 removal at constant gas flow 212 m^3/s (450,000 scfm) Mohave plant.[57]

efficiency was 82 percent. Successive stages removed more SO_2, up to about 99 percent at five stages. These tests were run at 212 m^3/s (450,000 scfm) at an inlet SO_2 concentration of 220 ppm. The L/G was 2.7 l/m^3 (20 gpm/1000 scfm) per stage. Both configurations of the vertical module operating on limestone showed a similar relationship.

An attempt was made to maximize use of cooling tower blowdown as makeup to the scrubber system, although it contained 12,000 ppm total dissolved solids. The horizontal module used 75 percent cooling water blowdown without formation of scale in the scrubber. The tests on the TCA and packed grid (PPA) absorbers were inconclusive, although the PPA showed no signs of scale with 40-percent cooling tower blowdown. Details of the water balances are shown in Figure 3-8.

Water was lost from the horizontal system in the following ways:[58]

9.03	l/s	(148 gpm)	as vapor in the flue gas
0.24	l/s	(4 gpm)	evaporation from the pond
0.24	l/s	(4 gpm)	with the sludge
9.51	l/s	(156 gpm)	Total

7.14 l/s (117 gpm) of makeup was supplied by cooling tower blowdown and 2.4 l/s (39 gpm) as fresh water.

Water was lost from the vertical module in the following ways:

9.03	l/s	(148 gpm)	as vapor in the flue gas
0.40	l/s	(6.5 gpm)	as evaporation from the pond
0.34	l/s	(5.5 gpm)	with the sludge
9.77	l/s	(160 gpm)	Total

Makeup was supplied from two sources: 5.9 l/s (96 gpm) fresh water and 3.9 l/s (64 gpm) cooling tower blowdown.

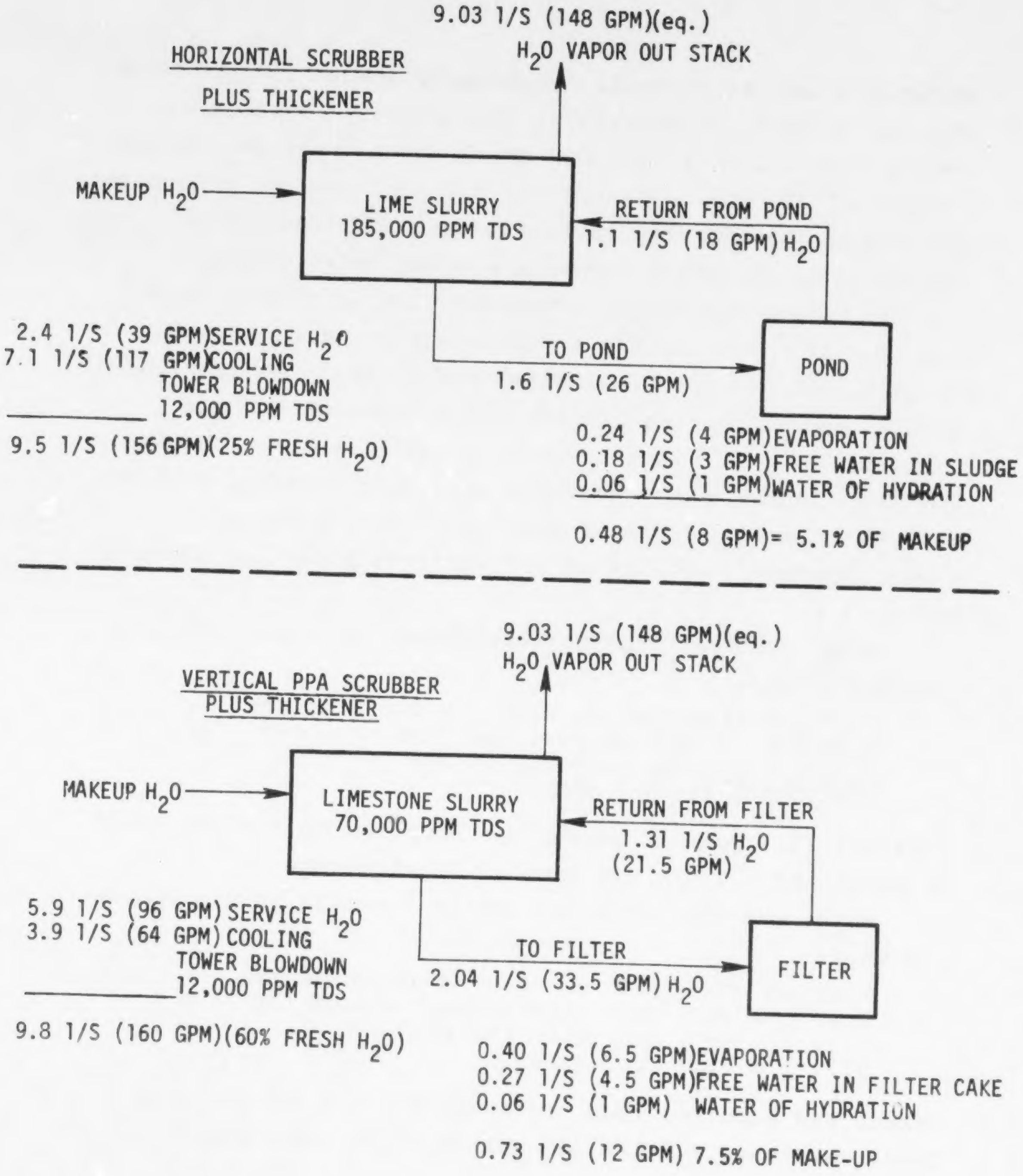

9.03 l/S (148 GPM)(eq.)
H_2O VAPOR OUT STACK
HORIZONTAL SCRUBBER
PLUS THICKENER
MAKEUP H_2O
LIME SLURRY
185,000 PPM TDS
RETURN FROM POND
1.1 l/S (18 GPM) H_2O
2.4 l/S (39 GPM) SERVICE H_2O
7.1 l/S (117 GPM) COOLING TOWER BLOWDOWN 12,000 PPM TDS
9.5 l/S (156 GPM)(25% FRESH H_2O)
TO POND
1.6 l/S (26 GPM)
POND
0.24 l/S (4 GPM) EVAPORATION
0.18 l/S (3 GPM) FREE WATER IN SLUDGE
0.06 l/S (1 GPM) WATER OF HYDRATION
0.48 l/S (8 GPM)= 5.1% OF MAKEUP
9.03 l/S (148 GPM)(eq.)
H_2O VAPOR OUT STACK
VERTICAL PPA SCRUBBER
PLUS THICKENER
MAKEUP H_2O
LIMESTONE SLURRY
70,000 PPM TDS
RETURN FROM FILTER
1.31 l/S H_2O
(21.5 GPM)
5.9 l/S (96 GPM) SERVICE H_2O
3.9 l/S (64 GPM) COOLING TOWER BLOWDOWN 12,000 PPM TDS
9.8 l/S (160 GPM)(60% FRESH H_2O)
TO FILTER
2.04 l/S (33.5 GPM) H_2O
FILTER
0.40 l/S (6.5 GPM) EVAPORATION
0.27 l/S (4.5 GPM) FREE WATER IN FILTER CAKE
0.06 l/S (1 GPM) WATER OF HYDRATION
0.73 l/S (12 GPM) 7.5% OF MAKE-UP

Figure 3-8. Water balances - Mohave plant.[59]

Operating History - Table 3-7 shows the performance history of the system. Since this was a test facility, several design changes were made during the period and this work contributed to the unavailability. It should be emphasized that the problems with the horizontal mist eliminator gas flow distribution, presented in Table 3-8, once solved, would not be a problem to future modules of this type. As a result, operability data are not very meaningful.

Several equipment-related test programs were conducted. For instance, two types of mist eliminators were tried, one for the horizontal module and one for the vertical module (Figure 3-9). It was concluded that a horizontal-type mist eliminator with intermittent wash on the front and back sides was preferred to the two section vertical-type mist eliminator with the described wash configuration.[62] Also, two types of reheat systems were investigated (Figure 3-10). The same amount of heat input was used for both the direct- and the indirect-type reheater. The indirect reheat did not show corrosion attack because the steam eliminators were only in contact with ambient air; the direct-type did show corrosion attack.[63]

3.1.3 Description of Foreign Installations

About half the Japanese FGD installations use lime or limestone for the absorbent in the circulating slurry.[64] Table 3-9 gives performances of SO_2 scrubbers on large boilers in Japan using lime slurry. There are also many lime scrubbing systems on smaller boilers and on new combustion processes.[65] Of the 13 boilers, 10 have SO_2 removal efficiencies in excess of 90 percent, some as high as 98 percent. In all except one case, the boilers are firing oil of low- to medium-sulfur content. One boiler is firing medium-sulfur coal.

Table 3-7. UNAVAILABILITY HISTORY - MOHAVE 170 MW TEST MODULES PROGRAM[60]

JANUARY 16, 1974 TO FEBRUARY 9, 1975 HORIZONTAL	HOURS
1. MODIFY AND REPAIR PLASTIC DEMISTER BLADES	503
2. CORRECT BOOSTER FAN BALANCE PROBLEMS	317
3. REPAIR CRITICAL PUMPS	256
4. REPLACE WORN-OUT SPRAY NOZZLES	238
5. MODIFY INLET GAS FLOW DISTRIBUTION	162
6. REPAIR HOPPER LEAKS	135
7. REMOVE HARDHAT FROM THICKENER	82
8. MODIFY SLAKING WATER TO PREVENT SCALE	45
9. CONDUCT INSPECTIONS FOR LONG RUNS	19
TOTAL	1757
TOTAL CALENDAR TIME	9328

PERCENT UNAVAILABILITY = 18.7

NOVEMBER 2, 1974 TO JULY 2, 1975 VERTICAL	HOURS
1. REPAIR GRIDS AND REDISTRIBUTE TCA BALLS	710
2. CLEAN SCALE FROM SCRUBBER INTERNALS	344
3. REPAIR OR REPLACE PLUGGED NOZZLES	153
4. REPAIR LEAKS IN TRAP-OUT TRAY	120
5. REPAIR/REALIGN PPA PACKING	85
6. CORRECT BOOSTER FAN TRIP PROBLEMS	72
7. CONDUCT INSPECTIONS FOR LONG RUNS	55
8. REMOVE HARDHAT FROM THICKENER	46
TOTAL	1585
TOTAL CALENDAR TIME	5813

PERCENT UNAVAILABILITY = 27.2

Table 3-8. OPERATIONAL DATA - MOHAVE HORIZONTAL FGD UNIT

JANUARY 16, 1974 TO FEBRUARY 9, 1975[61]

Months	Hours in period	Available	FGD hours called upon	Operating	Boiler hours	Parameter			
						1	2	3	4
						Availability	Reliability	Operability	Utilization
1/74	360	319	358	317	358	89	89	89	88
2/74	672	402	481	393	481	60	82	82	58
3/74	744	592	590	501	682	80	85	73	67
4/75	720	716	563	559	616	99	99	91	78
5/74	744	695	620	571	704	93	92	81	77
6/74	720	552	695	552	695	77	79	79	77
7/74	744	468	596	468	596	63	79	79	63
8/74	744	744	615	615	615	100	100	100	83
9/75	720	720	575	375	647	100	100	58	52
10/75	744	506	540	398	540	68	74	74	53
11/74	720	333	624	249	708	46	40	35	35
12/74	744	739	303	298	572	99	98	52	40
1/75	744	688	591	515	636	90	87	81	69
2/75	208	116	208	116	208	56	56	56	56

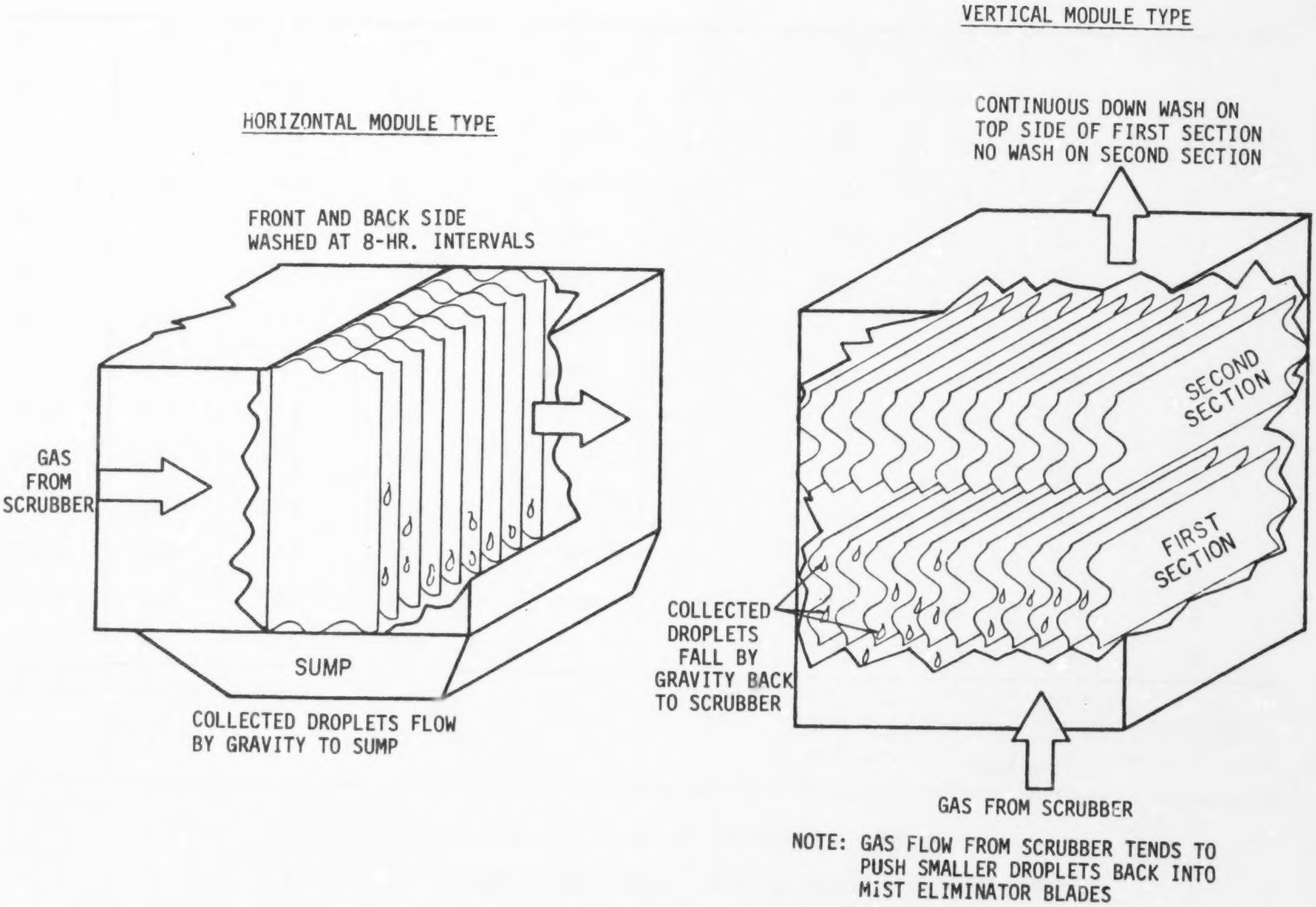

HORIZONTAL MODULE TYPE
FRONT AND BACK SIDE WASHED AT 8-HR. INTERVALS
GAS FROM SCRUBBER
SUMP
COLLECTED DROPLETS FLOW BY GRAVITY TO SUMP
VERTICAL MODULE TYPE
CONTINUOUS DOWN WASH ON TOP SIDE OF FIRST SECTION NO WASH ON SECOND SECTION
SECOND SECTION
FIRST SECTION
COLLECTED DROPLETS FALL BY GRAVITY BACK TO SCRUBBER
GAS FROM SCRUBBER
NOTE: GAS FLOW FROM SCRUBBER TENDS TO PUSH SMALLER DROPLETS BACK INTO MIST ELIMINATOR BLADES

Figure 3-9. Mist eliminators Mohave plant.[66]

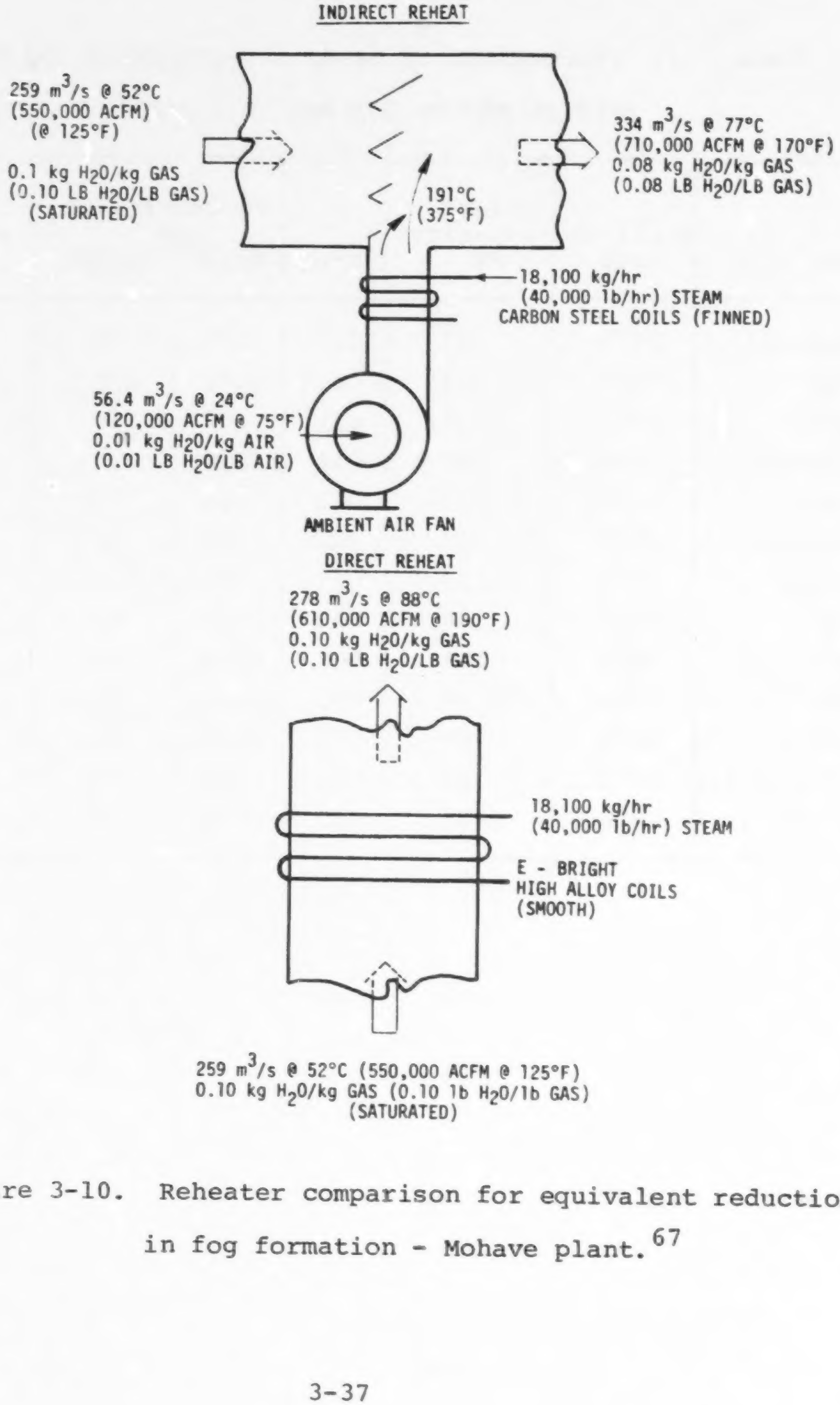


Figure 3-10. Reheater comparison for equivalent reduction in fog formation - Mohave plant.[67]

Table 3-9. PERFORMANCE OF LARGE SO_2 SCRUBBERS USING LIME SLURRY ON BOILERS IN JAPAN[68]

Plant site	Start-up date	Plant capacity, MW	Fuel	SO_2 concentration, ppm		Removal efficiency, %
				Inlet	Outlet	
Amagasaki	1972	32	Oil	700	70	90
Omuta	1972	164	Coal	2000	200	90
Kainan	1974	150	Oil	550	60	89
Hachinohe	1974	122	Oil	850	85	90
Karita	1974	175	Oil	800	75	91
Amagasaki	1975	125	Oil	500	50	90
Niigata	1975	170	Oil	700	70	90
Ehime	1975	86	Oil	1700	60	96
Mizushima	1975	192	Oil	1050	40	96
Owase	1976	2 x 375	Oil	1500	35	98
Iwakuni	1976	64	Oil	1400	50	96
Confidential	1976	125	Oil	500	65	87
Confidential	1976	384	Oil	550	65	88

All of the Japanese systems have attained an operability* of 95 percent or better. The Omuta Plant, firing medium-sulfur coal, for example, has been trouble-free since start-up in 1972. Another example is the Karita plant, treating flue gas from an oil-fired boiler. It went into operation in November 1974 and operated at 100-percent availability through 1976 (last report date).[69] Such high operability is normal on Japanese systems. Several reasons are given as to why Japanese operability is so much better than that in the United States:[70]

- Japanese emission legislation and enforcement is much stricter than in the United States. As a result, systems are installed with no bypass and must operate or the boiler cannot operate. This puts increased emphasis on installing reliable FGD systems.
- The Japanese systems are permitted to operate open-loop and to have a purge stream to remove soluble ions from the system.
- The Japanese systems tend to operate at a lower stoichiometric ratio, which tends to reduce the solid loading and also the plugging in the absorber.
- Japanese systems tend to be better designed, with presaturators, vertical mist eliminators, and low-sulfur oil reheat systems. Most systems in the United States have not installed these items for reasons of cost.

Oil-fired boilers tend to operate at slightly lower excess air rates and produce less fly ash compared with coal-fired boilers. However with high-efficiency fly ash removal preceeding an FGD system, and design for the greater excess air levels, lime systems should work equally well on coal fuel.

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* Operability here equals an FGD plant's service time divided by the scheduled operating hours of the process.

Technology Advancements - Mist Eliminators - Because of the strict particulate emission regulations, the Japanese systems must collect all the mist after scrubbing. Much more efficient mist elimination systems have been used; they consist of vertical configurations with high gas velocities 6.1 m/s (20 fps). This high gas velocity usually requires special eliminator vane design to prevent reentrainment.[71]

Reheat - Japanese systems are usually installed with external oil-fired reheat. This system is usually more expensive to operate than in-line steam coils used in the United States, but has proved much more reliable.[72]

Scaling - Most Japanese units operate on low-sulfur fuel. This may account for the fact that most of the systems do not have major scale problems. When close to the stoichiometric amount (0.95 to 1.05) of lime is used, scaling is reduced when compared with a stoichiometric amount of 1.1 to 1.3. Most Japanese systems use 0.95 to 1 of the stoichiometric amount of lime required. Another factor may be the Japanese systems oxidize the recirculating slurry to sulfates and maintain smooth scrubber internals to reduce buildup; these factors, however, have not been well-documented.[73]

3.1.4 Engineering Design Parameters

Introduction - Sulfur dioxide removal efficiencies in excess of 95 percent have been achieved using lime slurry scrubbing on boilers in Japan. Although full-scale lime scrubbing units in this country have not been designed for the same high removal efficiencies (because the SO_2 regulations are not as strict as in Japan), test facilities in the United States have demonstrated these same high removal rates. These data are given in this section. Lime systems have successfully operated on both high- and low-sulfur coal applications. Attainment of these high removal efficiencies and successful operation is dependent on careful design and operation. The importance of key design and operating parameters is discussed in this section.

3.1.4.1 $\underline{SO_2 \text{ Removal Potential of the System}}$ - Test facilities in the United States have produced data that indicate SO_2 outlet concentrations from FGD scrubbing systems can be as low as 1 to 5 ppm with removal efficiencies approaching 100 percent. Results of some of these tests are summarized below:

- At the Mohave plant of South California Edison, a 170-MW test facility, SO_2 outlet concentrations were less than 5 ppm with SO_2 inlet concentrations of around 220 ppm. Net SO_2 removal efficiency was about 98 percent. These tests were conducted over a one-year period on low-sulfur coal.[74]

- Recent tests at the Paddy's Run plant of Louisville Gas and Electric have shown SO_2 removal efficiencies in excess of 99 percent on 3-percent-sulfur coal. This extremely high removal efficiency was due to the addition of magnesium oxide to the scrubbing slurry.[75] Tests run for a 4-week period in May and June 1976 resulted in 98-percent SO_2 removal.[76]

- The TVA Shawnee plant reported SO_2 removal efficiencies of 95 to 99 percent for tests from September 21, 1974, to November 11, 1974. The SO_2 inlet concentration ranged from 1800 to 3800 ppm during this time.[77]

Full-scale facilities in Japan have shown similar high performance, as documented in Section 3.1.3 of this report. At these facilities, outlet concentrations as low as 35 ppm with SO_2 removal efficiencies in excess of 95 percent have been attained.

In summary, current technology is capable of controlling SO_2 emissions to overall efficiencies in excess of 98 percent on both low- and high-sulfur coal applications, as shown at the test facilities described above.

The ability of an SO_2 scrubbing system to remove high levels of SO_2 is dependent on many design factors. The operating conditions and results at the above plants vary widely, but several key design parameters have proved important to high SO_2 removal efficiency. These key design variables include:

- L/G ratio - the mass transfer potential of the system is directly related to the ratio of liquid volume to gas volume. Sulfur dioxide removal efficiency is greatly increased by increasing L/G.
- pH - SO_2 removal efficiency is directly related to slurry pH.
- Magnesium ion concentration - recent tests of lime slurry systems have shown increases in SO_2 removal efficiency when magnesium is added to the slurry.
- Gas velocity - the liquid-gas contact time is directly related to the gas velocity through the absorber and thus the SO_2 removal efficiency is also directly effected by the gas velocity.
- SO_2 inlet concentration - if all other parameters remain constant, increased SO_2 loading in the inlet gas will reduce the overall system efficiency.

3.1.4.2 L/G Ratio - The L/G ratio is the flow of slurry recirculating through the absorber in l/s (gpm), divided by the flow of flue gas in m^3 (1000 ft^3/min).

The theoretical effect of L/G on SO_2 removal efficiency is shown in Figure 3-11 for a spray tower. This figure shows SO_2 removal efficiencies predicted by Bechtel's mathematical model for the Shawnee plant, based on actual test data.

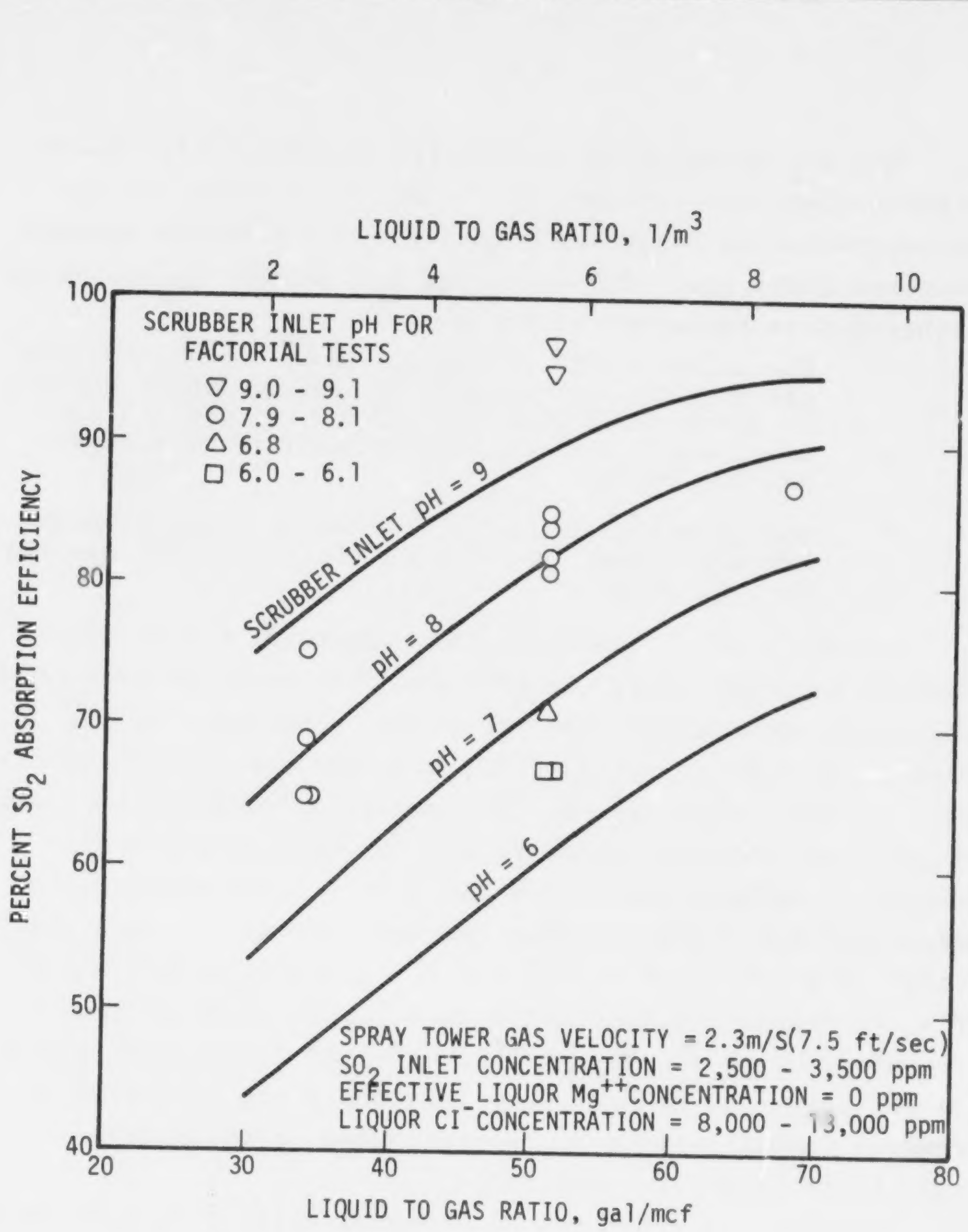

Figure 3-11. Liquid-to-gas ratio and scrubber inlet pH versus predicted and measured SO_2 removal, spray tower with lime, Shawnee plant.[78]

The SO_2 inlet (2500 to 3000 ppm) is for a high-sulfur eastern coal; gas velocity is 2.3 m/s (7.5 ft/s); Mg ion concentration is 0 ppm and chloride content varied between 8000 and 13000 ppm. The increased SO_2 removal exhibited by higher L/G is explained as follows:

- The increased flow creates better conditions for gas/liquid mass transfer at the interface because of increased turbulence and better gas distribution, which thereby increases the mass transfer driving force.
- The increased liquid rate creates a larger mass transfer area. This is multiplied by the greater driving force.

Figure 3-12, illustrates the effects of L/G on SO_2 removal in a TCA unit. In this example, both the predicted and measured data indicate the strong dependence of SO_2 removal on the L/G ratio. This test was run on high-sulfur coal at the Shawnee plant. The magnesium concentration was 0 ppm. The scrubber inlet pH was 7.9 to 8.1; and total height of spheres was 0.38 m (15.0 in.). The results are expressed for three scrubber gas velocities: 3.8 m/s (12.5 ft/s); 3.2 m/s (10.4 ft/s); and 2.5 m/s (8.3 ft/s). In a TCA, increased L/G does not significantly increase mass transfer, since the mass contact area is determined largely by the height of spheres. In this case, the increased SO_2 removal demonstrated must be attributed to the better gas/liquid interface conditions.

Figure 3-13 shows the SO_2 removal efficiency as a function of L/G in the horizontal test module at the 170-MW Mohave facility. The L/G shown is per stage in a four-stage system, so the total L/G is actually four times the indicated value. This facility fired low-sulfur coal (0.38 percent). The extremely high SO_2 removal efficiencies shown (98%)

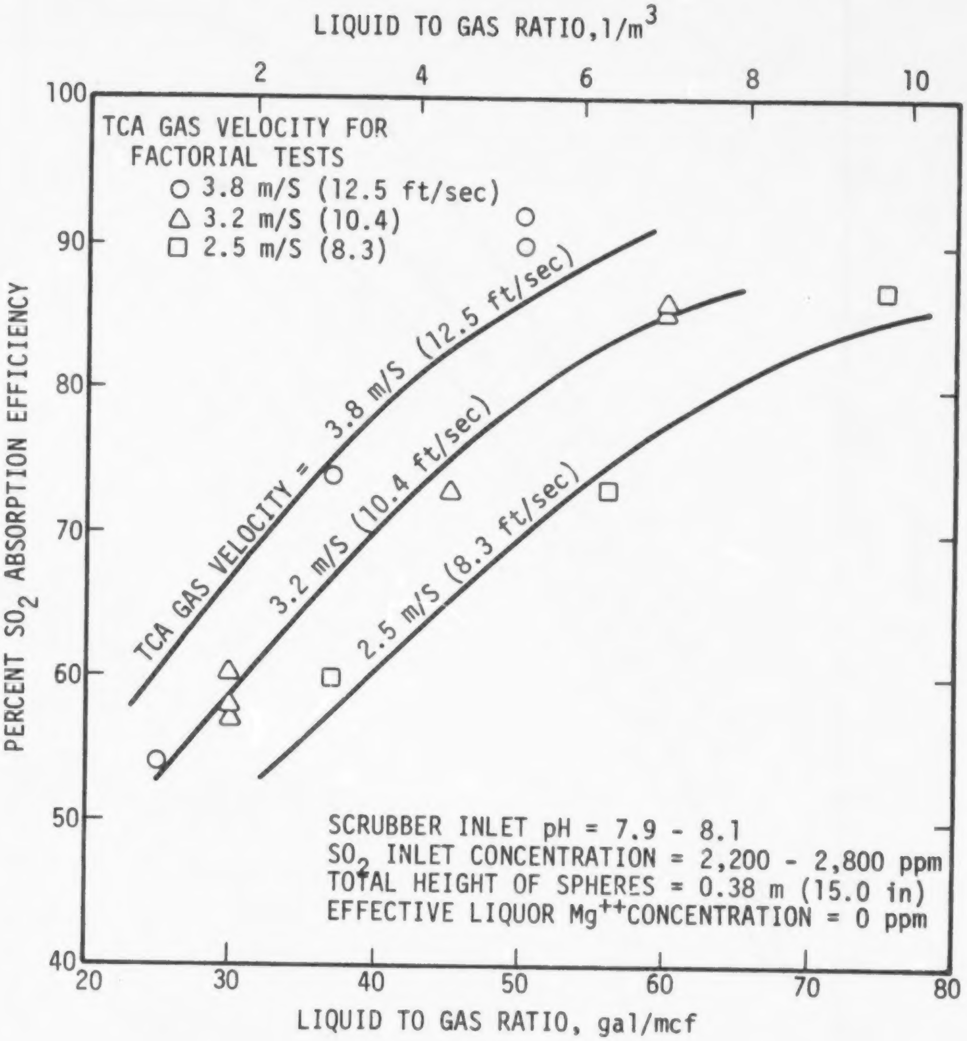

Figure 3-12. Liquid-to-gas ratio and scrubber gas velocity versus predicted and measured SO_2 removal, TCA with lime, Shawnee plant.[79]

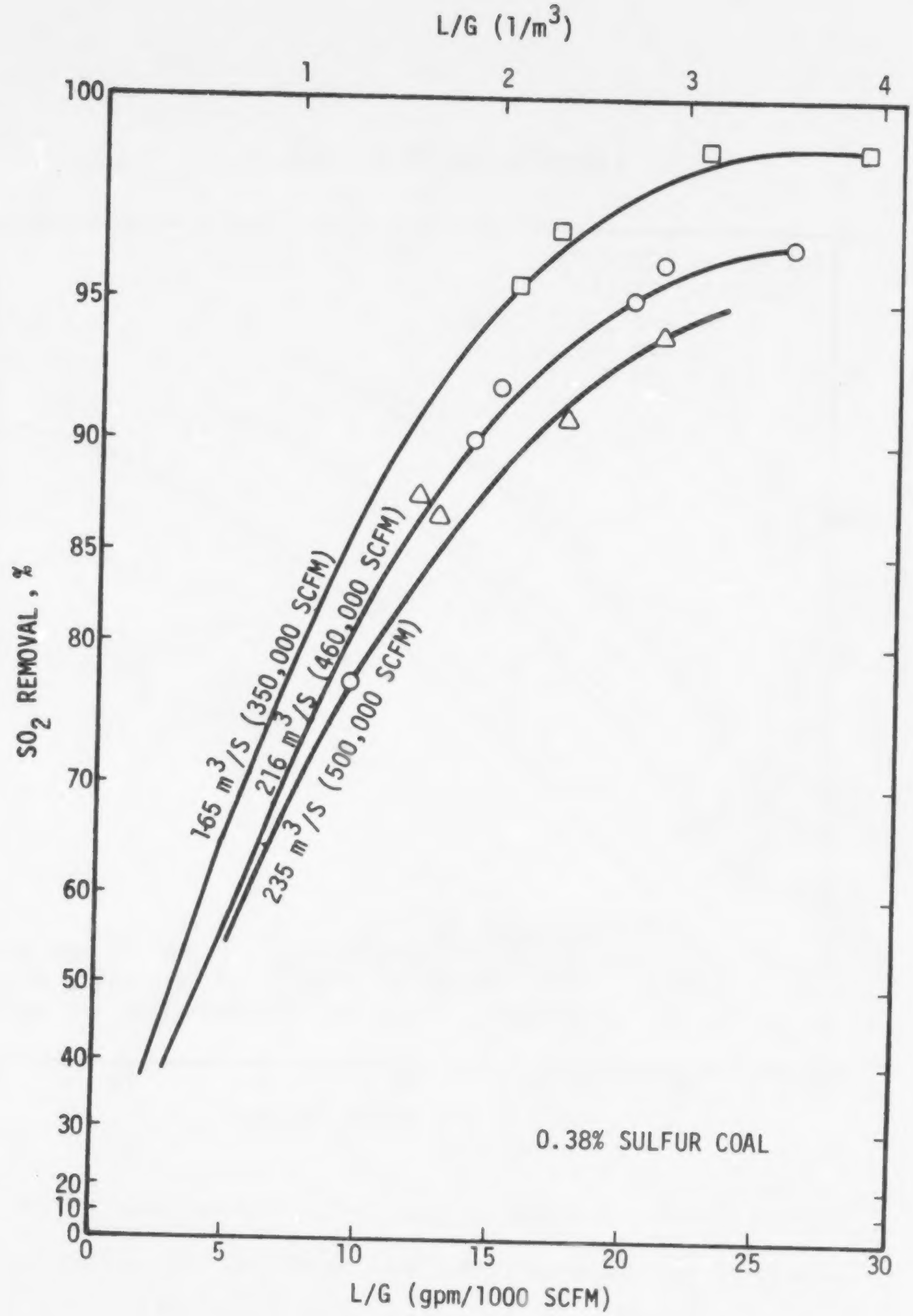

Figure 3-13. SO$_2$ removal versus L/G ratio, 170-MW horizontal module, Mohave plant.[80]

yielded SO_2 outlet concentrations less than 10 ppm. Three curves are shown for 235, 216, and 165 m^3/s (500,000, 460,000 and 350,000 scfm). As expected, SO_2 removal efficiencies were lower at the higher gas flow rates because of the shorter residence time. All three gas flow curves illustrate increased SO_2 removal efficiency at increased L/G. Since this is a spray tower, the increased SO_2 removal efficiency can be attributed to both increased contact area (which is directly proportional to liquid rate) and better gas/liquid interface characteristics.

3.1.4.3 pH - Johnstone, in research in 1935, found that the equilibrium vapor pressure of SO_2 over lime slurry was inversely proportional to the slurry pH (Figure 3-14). This phenomenon is caused by the higher alkalinity, which results in lower SO_2 equilibrium vapor pressure at higher pH.

The net effect of the pH on SO_2 removal efficiency for the TCA at the 10-MW Shawnee test facility is shown in Figure 3-15. The trend predicted by the Bechtel Model is indicated by the solid lines; the actual data points are also shown. Although the pH effect is not as graphic as the one shown primarily for L/G, the trend is obvious. Higher pH results in higher SO_2 removal efficiencies. When used in conjunction with a moderate L/G 8.04 l/m^3 (60 gal/1000 ft^3), high pH can achieve 90-percent removal (top curve in Figure 3-15). There are, however, limitations. Higher pH requires addition of excess lime. The excess lime, as well as raising the pH, will travel through the scrubber system as excess solids, and tend to increase scrubber scale formation. The higher SO_2 removal efficiency at higher pH is the result of increased alkalinity of the circulating slurry. As shown by the Johnstone graph, this tends to drive the SO_2 into the circulating slurry. The same relationship of

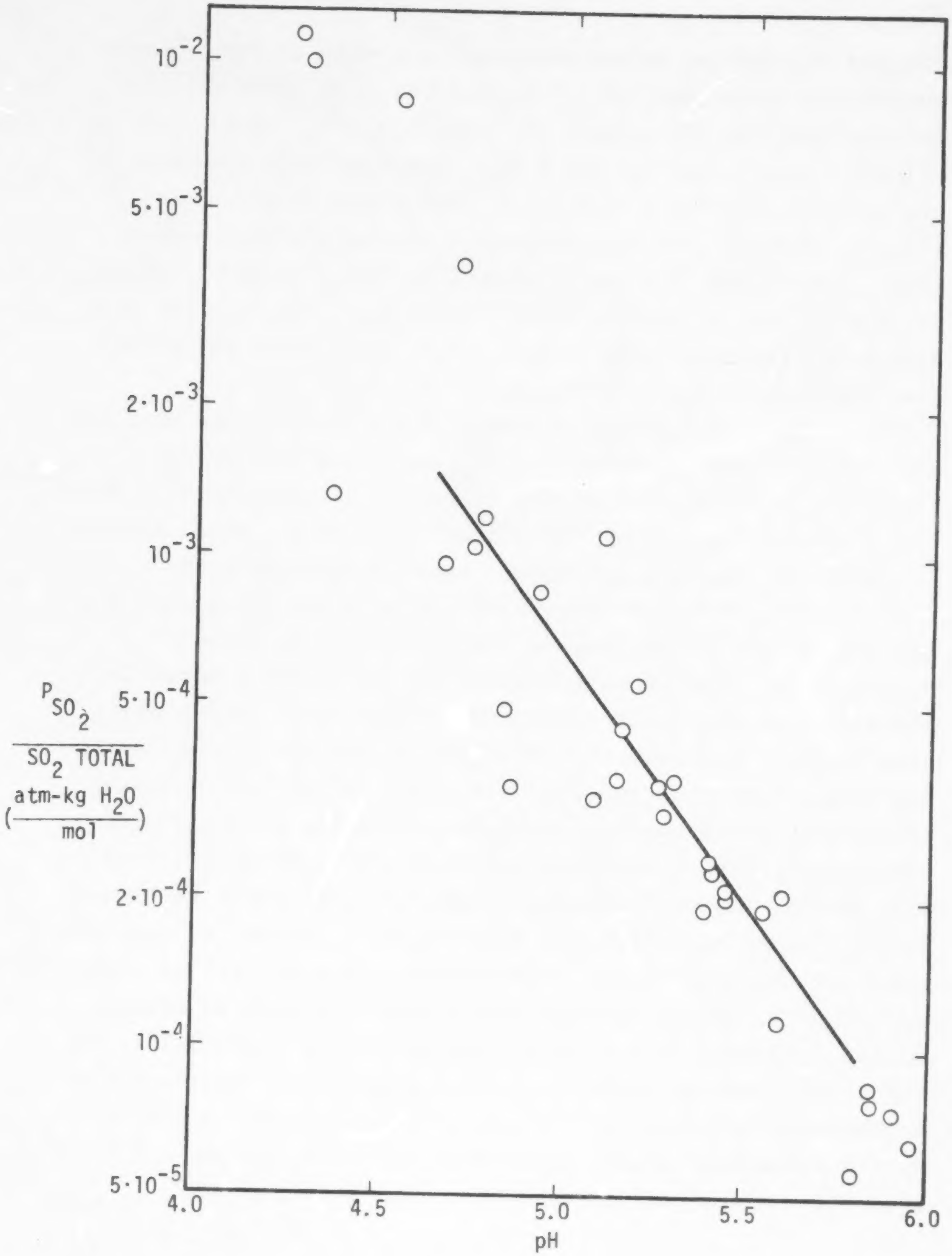

Figure 3-14. Effect of pH on SO_2 vapor pressure over buffered solutions.[81]

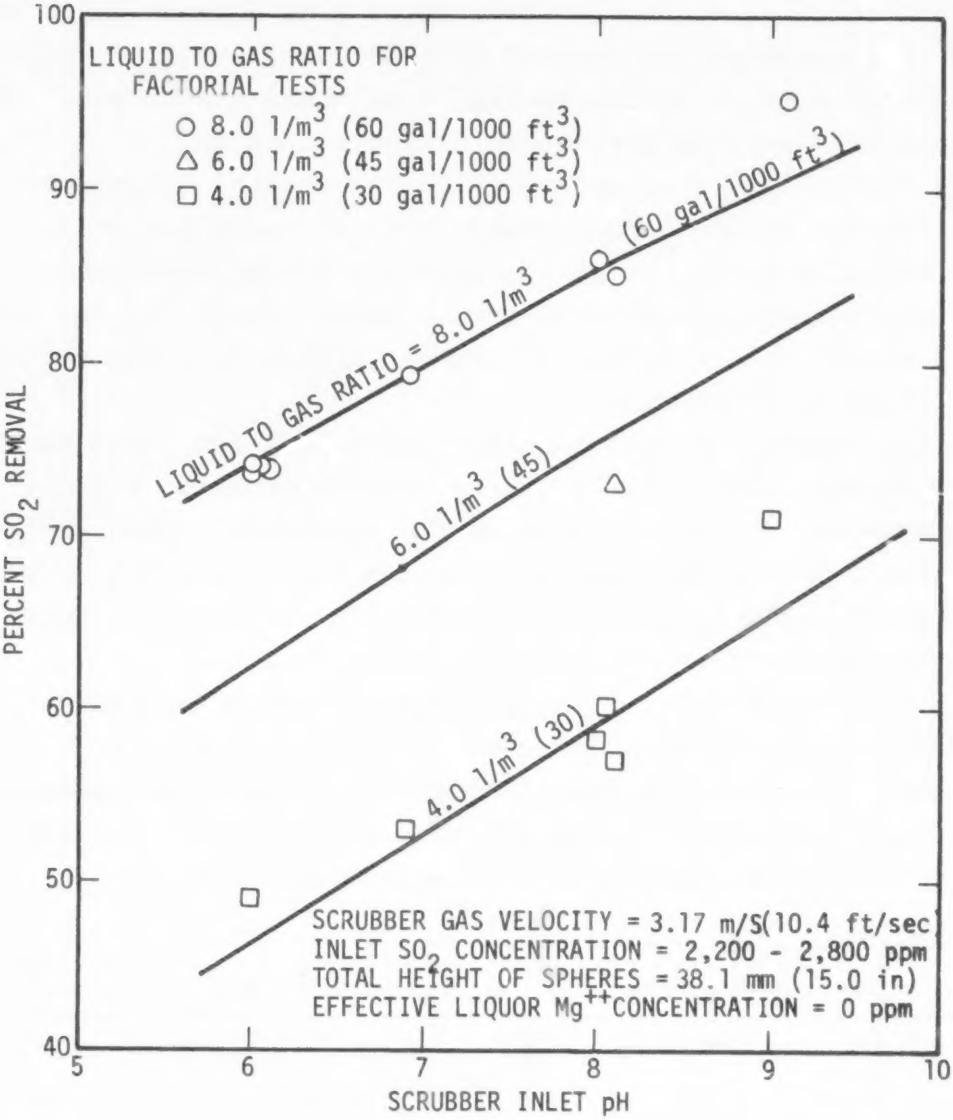

Figure 3-15. Scrubber inlet pH and liquid-to-gas ratio versus predicted and measured SO_2 removal, TCA with lime, Shawnee plant.[82]

higher SO_2 removal efficiencies at higher pH is expected for all types of absorber modules. Figure 3-16 shows the distribution of aqueous sulfite species as a function of pH. The relationships are derived from the disassociation constants of H_2SO_3. It is apparent from these graphs that the predominant sulfite species is a function of pH.

The predicted effect of pH on SO_2 removal efficiency for the TVA Shawnee spray tower is shown in Figure 3-17. The actual data do not agree closely with the predicted values, but the trend is apparent, particularly for the L/G of 6.8 l/m^3 (51 gal/1000 ft^3) curve, higher pH gives better SO_2 removal efficiencies.

3.1.4.4 Magnesium Ion Concentration - Addition of magnesium ions to the recirculating slurry has demonstrated superior SO_2 removal. Figure 3-18 shows the effect of magnesium ion concentration on SO_2 removal efficiency at the 0.564-m^3/s (1200-cfm) pilot plant run by Combustion Engineering Power Systems for medium- and high-SO_2 inlet loadings.

Full-scale units have also reported increased SO_2 removal with magnesium ion addition. For example, SO_2 removal efficiency at the Paddy's Run facility was measured with and without magnesium addition. Efficiency averaged 83 percent without magnesium; with magnesium, however, the following results were obtained:[85]

Date	Inlet SO_2 concentration, ppm	Outlet SO_2 concentration, ppm	Percentage removal
7/6/77	2150	1 to 5	99.7 to 99.9
7/8/77	2230	1 to 5	99.7 to 99.9

In addition, the Phillips Station conducted tests to demonstrate the effect of increased magnesium ion concentration. At this station the overall SO_2 removal efficiency increased

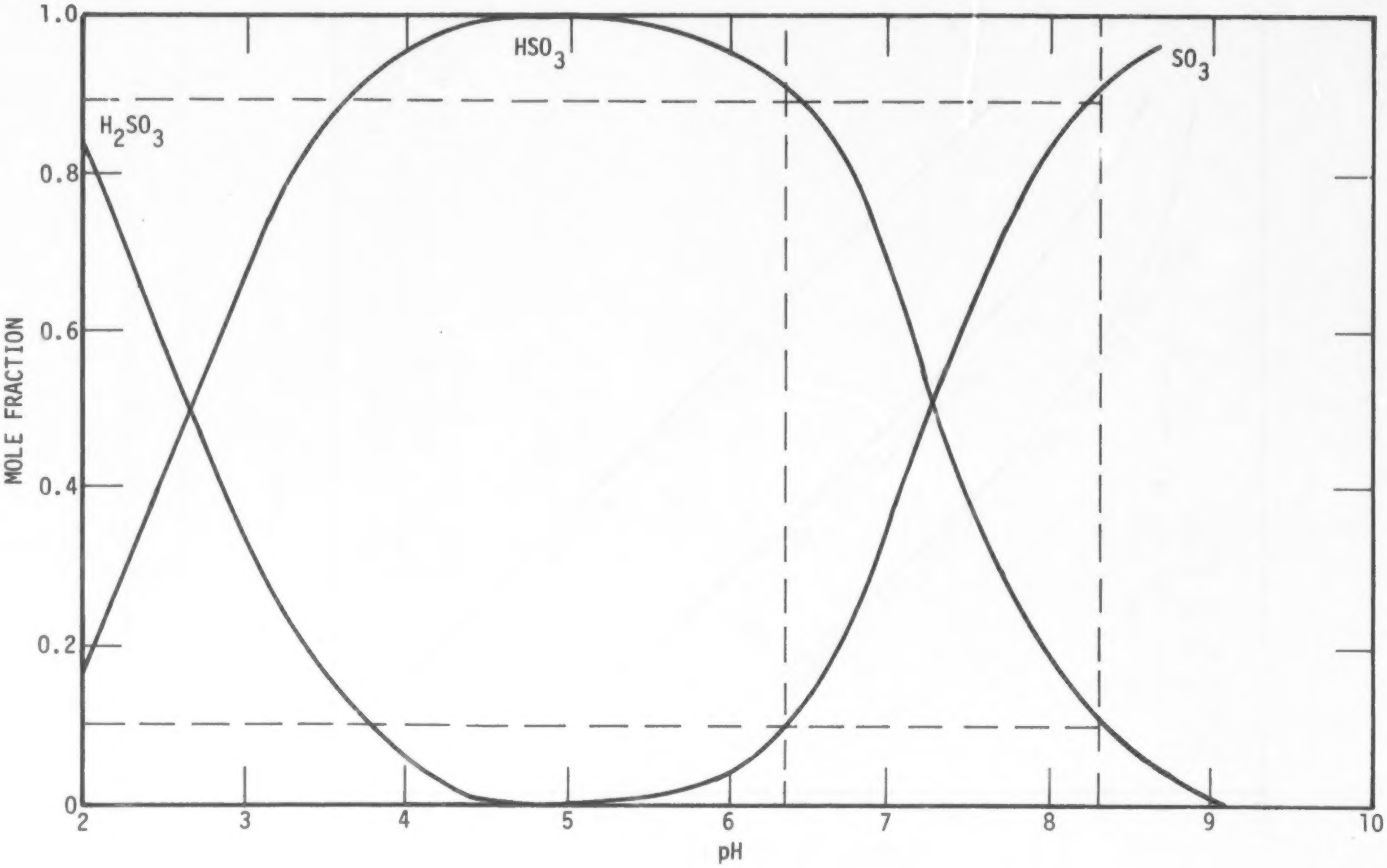

Figure 3-16. Distribution of aqueous sulfite species as a function of pH.[83]

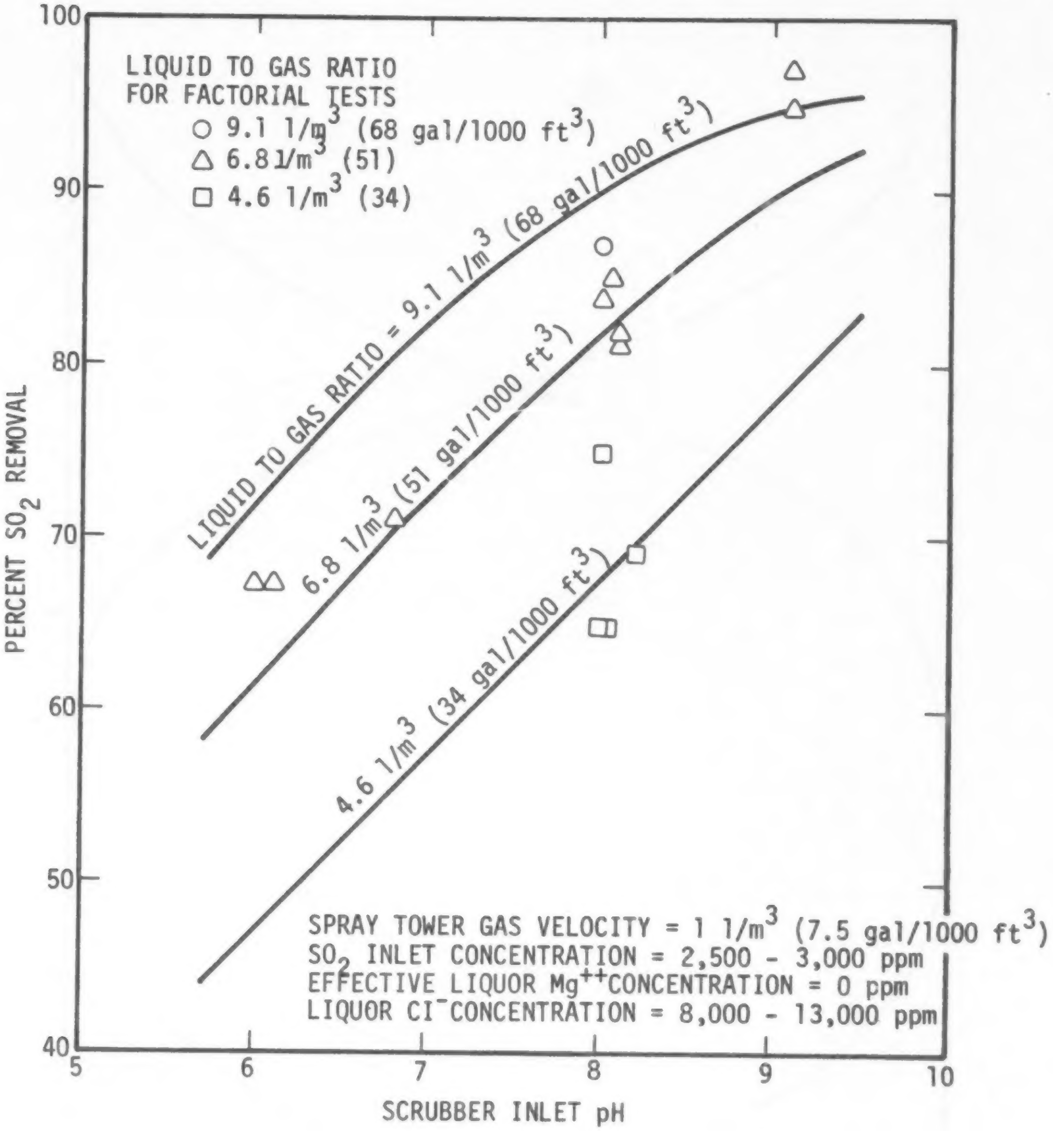

Figure 3-17. Scrubber inlet pH and liquid to gas ratio versus predicted and measured SO_2 removal, spray tower with lime, Shawnee plant.[84]

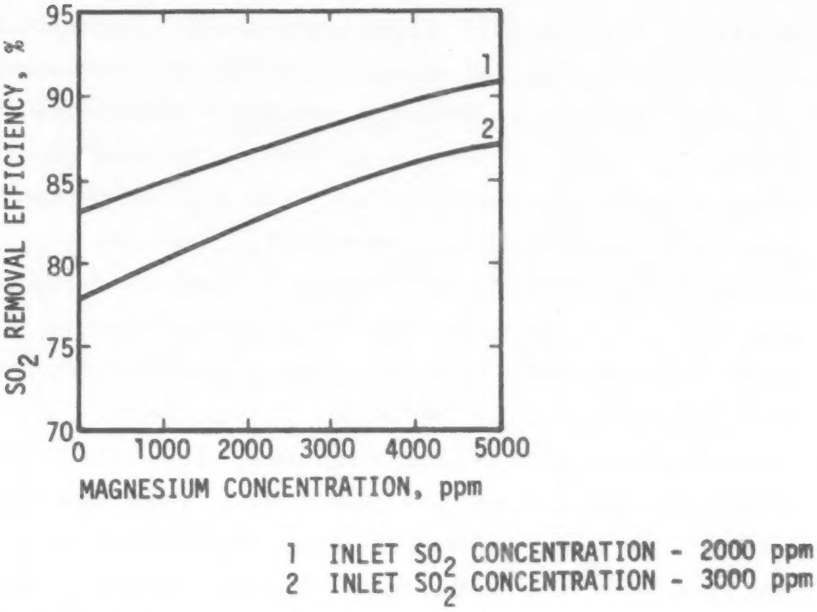

Figure 3-18. Effect of magnesium on SO_2 removal efficiency.[87]

from around 50-percent SO_2 removal to about 83-percent SO_2 removal when lime containing up to 10 percent magnesium ion was used.[86]

The reason for this phenomenon is that the magnesium ionic species are much more soluble than the calcium ionic species. Thus, the magnesium ions are in solution and ready to react with the bisulfite and sulfite ionic species. This increases the rate at which SO_2 can be removed from the system. Figure 3-19 illustrates this increased alkalinity as the magnesium ion concentration is increased.

3.1.4.5 Number of Contact Stages - When lime slurry contacts SO_2-laden flue gases, SO_2 is removed until the solution approaches equilibrium. As this happens, the solution has less potential to remove SO_2, and the SO_2 removal rate is therefore lower. One method used to maintain high SO_2 removal is to contact the SO_2-laden flue gas with fresh slurry in several stages. The effect of increased stages at the Mohave plant is shown in Figure 3-20. This case is for a horizontal cross-flow scrubber; additional stages are added to the end of the absorber. This can also be accomplished in the TCA scrubber by additional trays, or in a spray tower with additional spray modules. The figure shown is for low-sulfur coal with L/G of 2.7 l/m^3 (20 gal/1000 ft^3) per stage. As a result, the total L/G is the number of stages multiplied by 2.6 l/m^3 (20 gal/1000 ft^3).

3.1.4.6 Gas Velocity - The length of contact time between the slurry and the flue gas will influence the amount of SO_2 removal. The longer the slurry is in contact with the flue gas, the higher the overall SO_2 removal will be. This was demonstrated in the test facilities at Mohave shown earlier in Figure 3-13 for the horizontal module. The three flue gas rates are directly proportional to velocity, and the

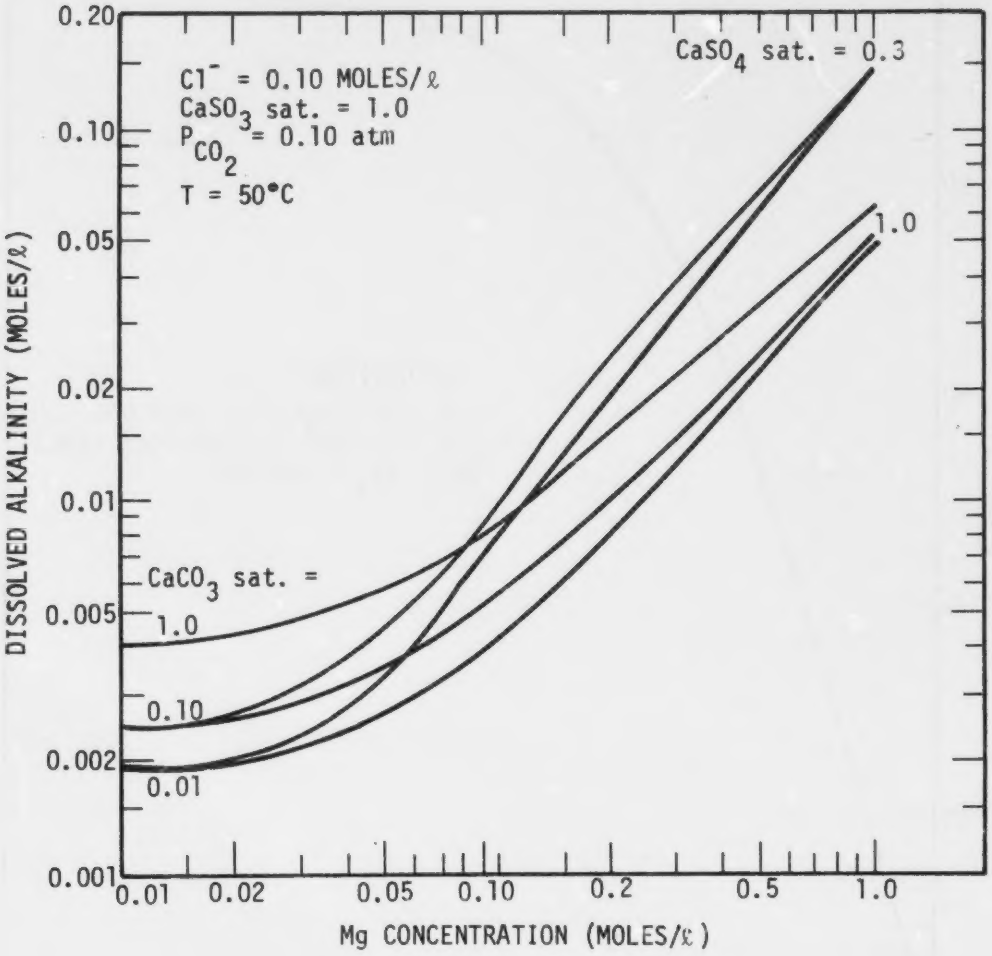

Figure 3-19. Dissolved alkalinity generated by addition of MgO.[88]

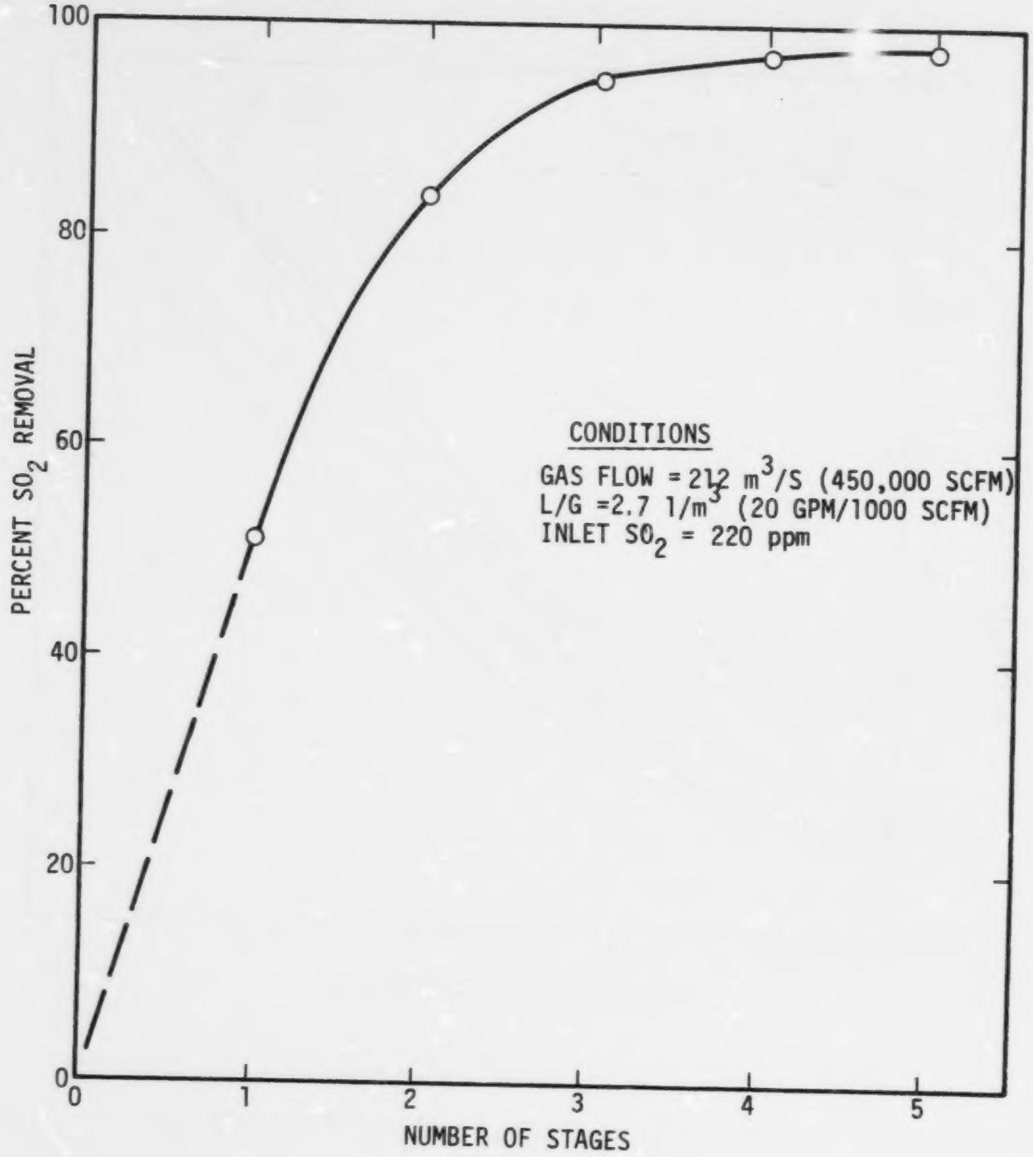

Figure 3-20. 170-MW horizontal SO_2 removal versus number of stages, Mohave plant.[89]

data show that the higher gas flow rates yield a lower SO_2 removal efficiency.

Figure 3-21, for the spray tower at the 10-MW Shawnee test facility, also shows the flue gas flow velocity inversely related to the SO_2 removal efficiency at a given liquid rate. The solid lines indicate the predicted values; individual data points are also shown. For a packed or turbulent type absorber, gas velocity is not so critical since increased agitation and mixing partially offset the decreased L/G ratio.

3.1.4.7 $\underline{SO_2\text{ Inlet Concentration}}$ - Higher SO_2 inlet loadings mean that less SO_2 is removed on a percentage basis at a fixed operating condition. For a given set of operating conditions, a lower SO_2 removal efficiency will be realized if the SO_2 inlet concentration is increased. This is illustrated in Figure 3-22, which is again based on Shawnee test results. These data show the theoretical and actual SO_2 removal efficiencies for both spray tower and TCA absorber, at an L/G of 11.4 l/m^3 (85 gal/1000 ft^3). Since all other SO_2 removal parameters remain fixed in this example, the SO_2 removal efficiency decreased as a function of inlet SO_2. By varying the operating conditions, for example by increasing the L/G ratio, a desired efficiency can be maintained.

3.1.4.8 Overall Effects - All the variables discussed above affect the system. To determine the overall effect, each would have to be analyzed in conjunction with the remaining factors. Bechtel has developed a model to define these effects for the 10-MW test facility at Shawnee. This model is useful for selecting the parameter needed to achieve high SO_2 collection efficiencies.[94]

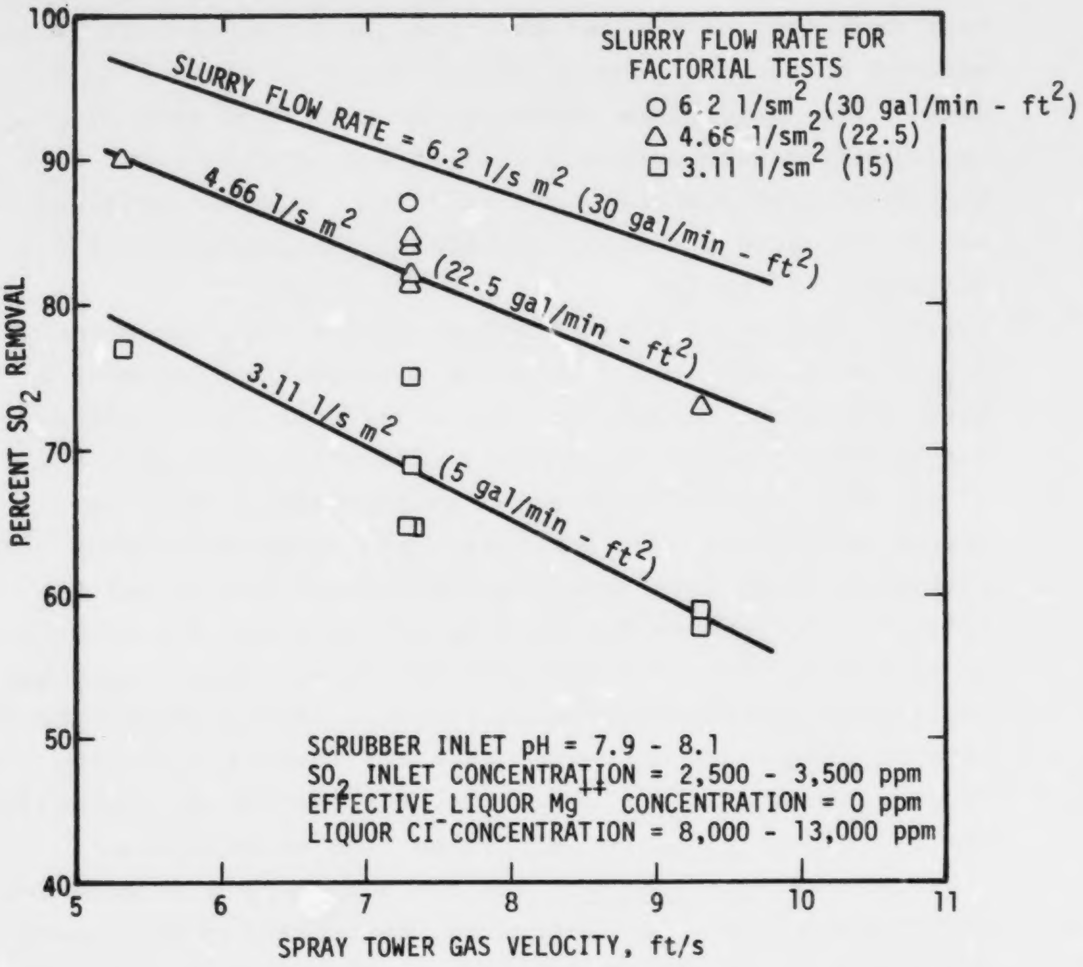

Figure 3-21. Gas velocity and slurry flow rate versus predicted and measured SO_2 removal, spray tower with lime, Shawnee plant.[90]

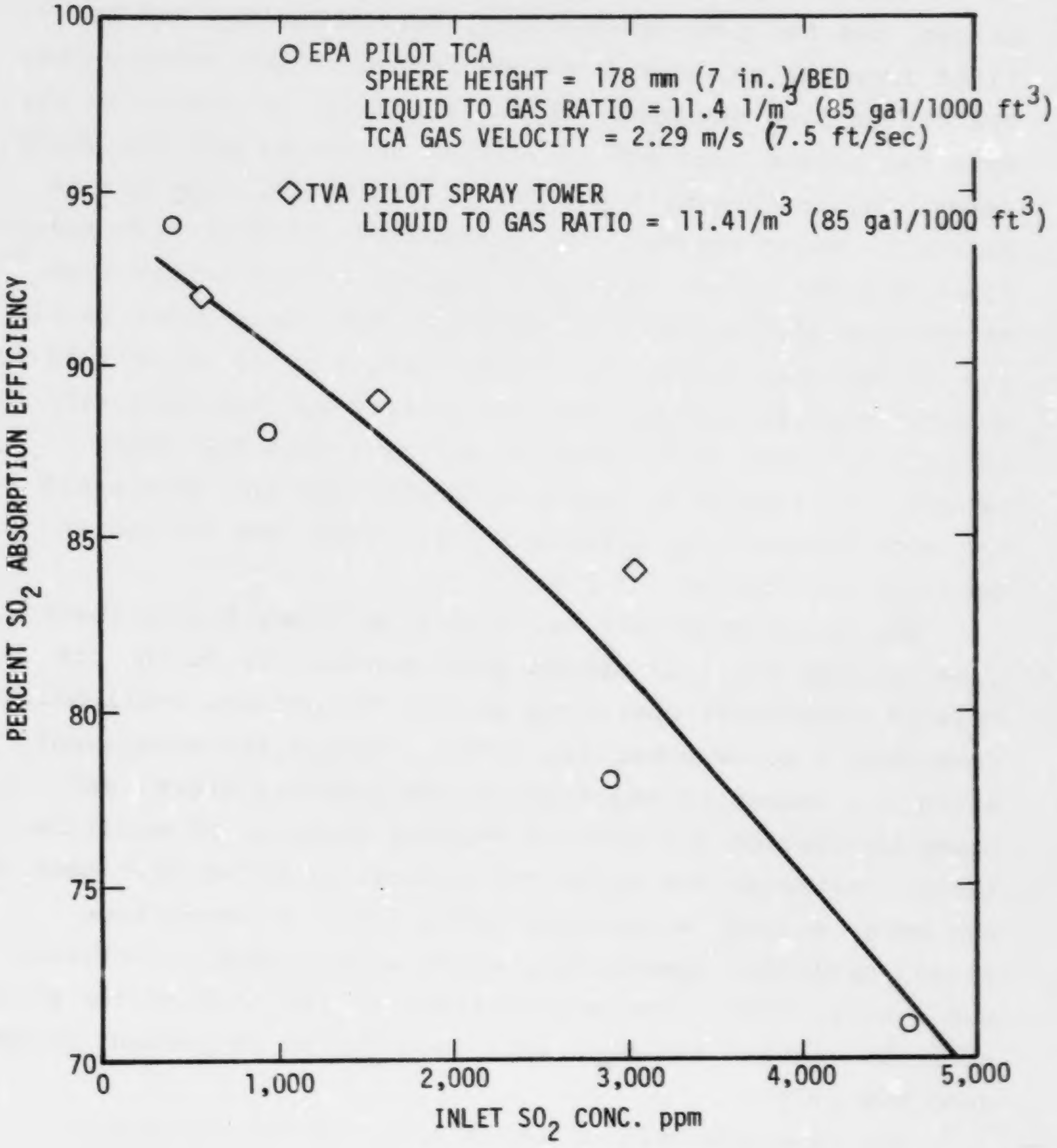

Figure 3-22. Effect of inlet SO_2 concentration on SO_2 absorption efficiency for fixed design condition.[91]

3.1.5 Operability

The Phillips lime scrubber, the first in the United States, has had poor operability, but technology has benefited from the experience gained from this and other units. The Elrama lime scrubber, which is similar in design to the Phillips system, was put in service approximately two years later. Although only a percent of the boiler flue gas is currently being treated, the operability of this five-module plant has been good, with only limited forced outage time.[92] Of the new generation, for instance, the Green River facility of Kentucky Utilities, which started up in January 1976, quickly rose to 100-percent operability and ran very well through February 1977, when it was shut down for stack repair. It started up again in April 1977 and once again has good operability (Figure 3-23). This information is detailed in Section 3.1.2.2.

The Bruce Mansfield No. 1 unit of Pennsylvania Power (see Section 3.1.2.2) showed good operability during its initial commercial operating period (98 percent availability from June 1 to December 31, 1976). During the subsequent months, a number of major operating problems arose, and power production was reduced pending repairs and modifications. Although the major FGD-related problems have been or are being solved, widespread stack liner failures have caused half-load operations, which will probably continue until early 1978. The availability of the half of the plant still in service has been very good (95 to 99 percent during June and July).

The Cane Run unit of Louisville Gas and Electric, as shown in Figure 3-24, initially showed good start-up operability, but was shut down in 1977 when river freezing prevented lime delivery. During initial operation,

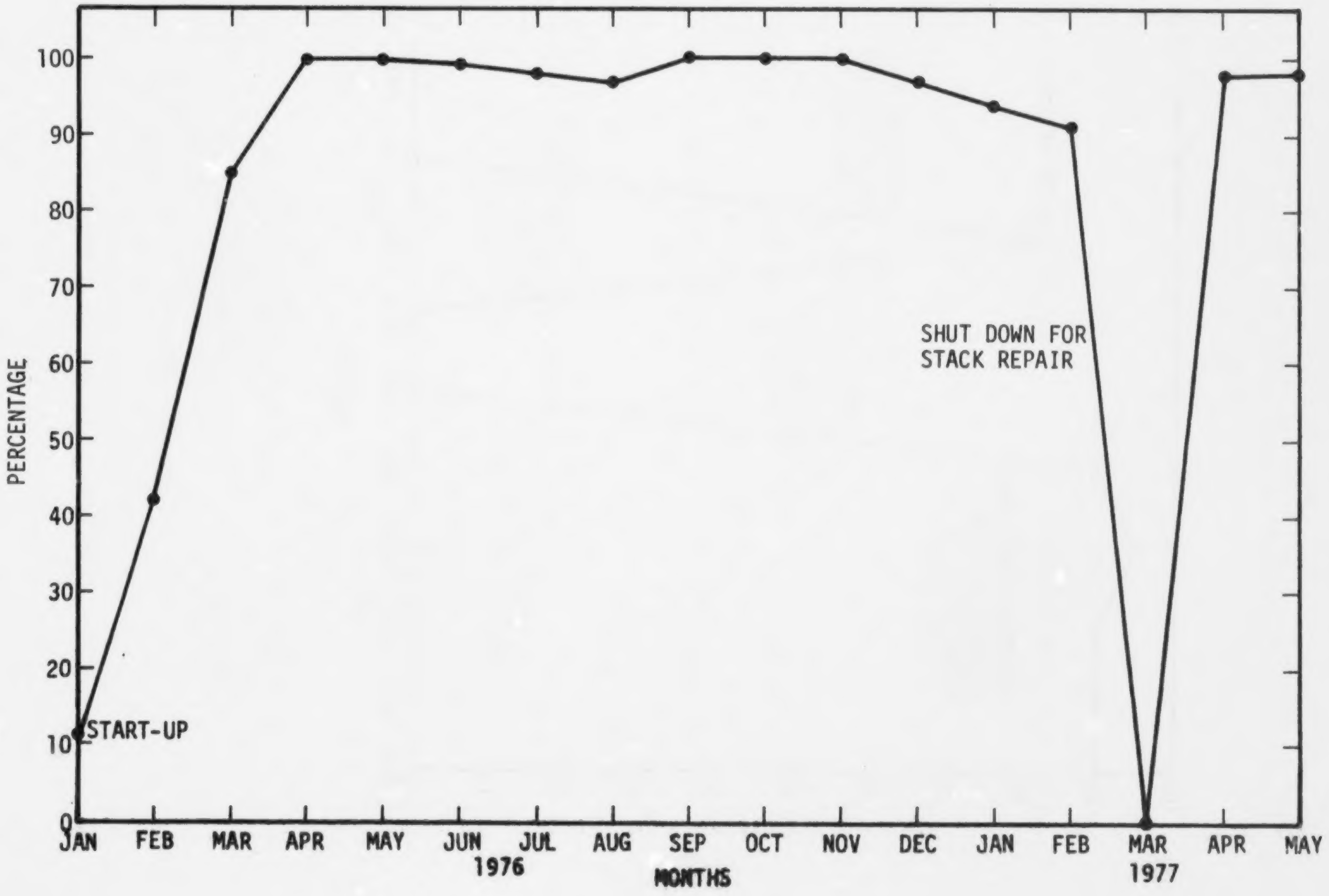

Figure 3-23. Scrubber system operability Green River No. 1, 2, and 3.[95]

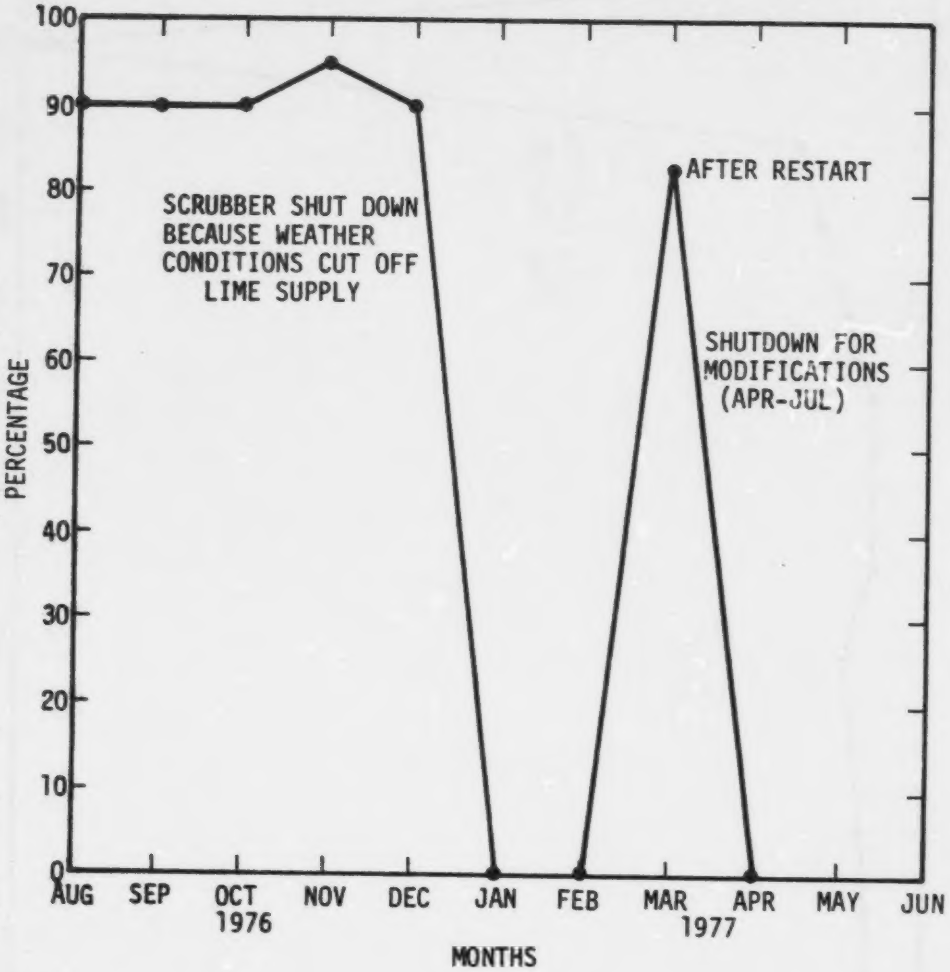

Figure 3-24. Scrubber operability,
Cane Run No. 4.[96]

several design errors were detected and the unit was shut down early in 1977 for adjustments; the mist eliminators were replaced with Chevron-type units; L/G was increased; an oil-fired reheater was added; and the stack lining replaced.

The Paddy's Run unit of Louisville Gas and Electric is an older unit and the boiler is used for peak loads only. As a result, the scrubber runs infrequently, but it does have good operability during its working periods (Figure 3-25).

The Colstrip No. 1 and No. 2 units of Montana Power have been operating since September 1975 and May 1976 respectively. Both are equipped with SO_2 and particulate scrubbing systems. The alkalinity of the fly ash and supplemental lime are the principal reagents. Systems availability for both units has been good and has steadily improved. For Unit No. 1, the annual average total availability values for 1976 and 1977 (August) were 86 and 89 percent respectively. Unit No. 2 values were 96 and 85 percent for 1976 (May startup) and 1977 (August) respectively.[93]

3.1.5.1 Effect of Increased SO_2 Removal on Operability

Numerous corrosion and erosion problems have attended the operation of lime scrubbing systems, as indicated in Section 3.1.2. As design techniques become more refined, however, corrosion problems are being solved with improved construction materials. For example, the recirculation pumps at the Phillips station were designed with Carpenter 20-type wetted parts. It was soon apparent that this material is not suited to lime slurry scrubbing. Subsequent installations have incorporated rubber-lined or hardened iron parts and have experienced few difficulties.[98] Corrosion problems do continue in some components. Linings on vessels and stacks,

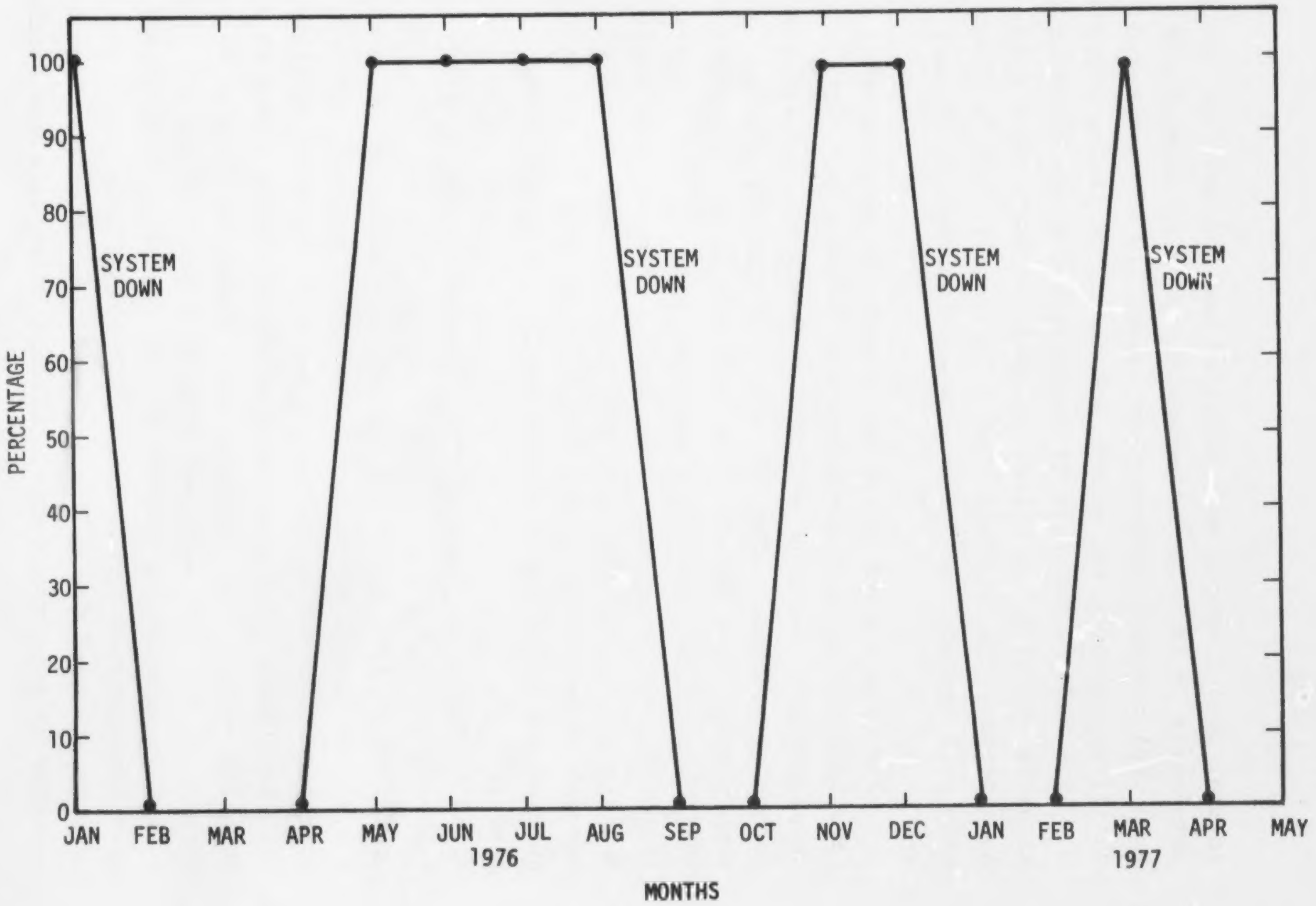

Figure 3-25. Scrubber operability - Paddy's Run No. 6.[97]

and in-line reheat coils have failed repeatedly. In some cases, however, unlined carbon steel reheaters have performed well. Although the higher reactivity of lime makes it difficult for localized volumes of acidic solution to accumulate, some degree of corrosive attach has occurred in dampers, turning vanes, and expansion joints. The solution to the problem is the replacement of corroded parts with materials capable of withstanding acidic corrosion, such as 316 L SS or Monel.

Stacks associated with lime scrubbers have shown a high degree of corrosion (at Cane Run and Green River for example) apparently because of the lack of reheat in these lime scrubber systems, or because of improper lining and mist elimination. Operators of these systems are planning to incorporate reheat.[99] Increased SO_2 removal should not amplify the corrosion problem; in fact, it should reduce the problem of producing less corrosive flue gas.

Scaling and plugging problems are still of major consequence in this country. Approaches to solving these problems include regular shutdown and cleaning of each absorber, addition of magnesium ion to the slurry, and better pH and solids control. The magnesium ion dissolved in the slurry increased alkalinity and the rate of chemical reactions. Both these techniques show some merit, although neither one alone provides the total answer to scaling and plugging problems.

At present, the operational systems generally use 1.1 to 1.3 times the amount of lime theoretically required. If continued, this strategy may increase chemical and mechanical problems caused directly by plugging and scaling, which in turn are due to the heavier load of solids in the system.

New technology that might over-come these problems is described in Section 3.1.2.4.

Scale and plugging also occur in the mist eliminator section of the system. Washing techniques have been developed that appear to ease the problem.

3.1.6 Vendor Guarantees

For systems currently going on-line, the vendors are guaranteeing higher SO_2 removal efficiencies. For example, the Conesville Station has an 87.6 percent vendor guarantee;[100] for the Cane Run Station it is 85 percent;[101] and at Bruce Mansfield 92 percent.[102] A brief vendor survey (see Section 4.2) was conducted for this review; all vendors surveyed would guarantee 90-percent or better SO_2 removal efficiency on their systems. In addition, vendors are now guaranteeing operability periods. The Green River and Conesville Generating Stations for example, had six-month operability guarantees. This means that the vendor will guarantee a certain operability, usually 90 percent for a given length of time.

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3.2 LIMESTONE SLURRY FLUE GAS DESULFURIZATION SYSTEMS

The limestone slurry scrubbing process is similar to the lime slurry process described in Section 3.1. Since the basic chemistry and SO_2 removal mechanisms are similar to the lime-based systems, technical details that were explained in Section 3.1 will not be repeated here. Instead, this section will highlight any differences in the two processes and discuss the performance results of existing limestone slurry FGD systems.

3.2.1 Process Description

Development of this process began much earlier than the lime scrubbing systems. More than 40 years ago, calcium carbonate ($CaCO_3$) was commercially used to remove SO_2 from waste gases. The process was applied to flue gases from power plants in England. Four power plants were equipped with scrubbers before 1940. Although these units were inefficient in comparison with those being built today, two were still in use in 1970.

Except for small-scale studies by TVA in the 1950's, little interest was reported in limestone FGD until after 1960, when research began almost simultaneously in Japan, the Soviet Union, and the United States. Work in the Soviet Union was directed primarily toward abatement of SO_2 pollution from the metallurgical industry; several systems are now in operation.[1] The Japanese now have 12 major (larger than 100 MW) limestone-based FGD systems on utility and industrial boilers.[2]

In the United States, the Universal Oil Products Company tested a mobile-bed limestone scrubber on a power plant in Wisconsin in 1965. The next year, Combustion Engineering followed with a pilot plant on a Detroit Edison power station. An acceleration of effort began shortly before 1970. Zurn Industries, in cooperation with the National Air Pollution Control Administration (now EPA), constructed a pilot limestone FGD system that was tested on an oil-fired generating station at Key West, Florida, and later on the coal-fired Shawnee station of the TVA at Paducah, Kentucky. Also under EPA contract Bechtel built and operates a large pilot plant at the Shawnee station and conducts tests of both lime and limestone scrubbing. During the EPA program, TVA modified one of the scrubbers for a test series with other types of absorber internals. Babcock and Wilcox performed bench-scale and pilot-plant testing of limestone FGD at their Alliance, Ohio, research and development center. Southern California Edison, in a joint venture with five other western Power utilities, tested limestone as a scrubbing reagent at the Mohave generating station in Nevada. Equipment was later moved to Four Corners, New Mexico, for additional testing. Combustion Engineering, in cooperation with Northern States Power Company, tested a prototype unit at the utility's Black Dog station. Detroit Edison and Peabody Engineered Systems conducted a pilot-plant program at the St. Clair power plant from 1971 to 1973. A full-scale unit was built at St. Clair in 1974 and was tested until 1976.[3]

A significant amount of theoretical work was conducted in support of the plant testing and equipment development. Most of the published findings are from studies funded under EPA contracts.

The study of calcium- and magnesium-based FGD systems has been one of the most intensive and broadly-based industrial research programs ever conducted in this country. A massive body of data has been accumulated in public record, and still more is held as proprietary information by various companies.

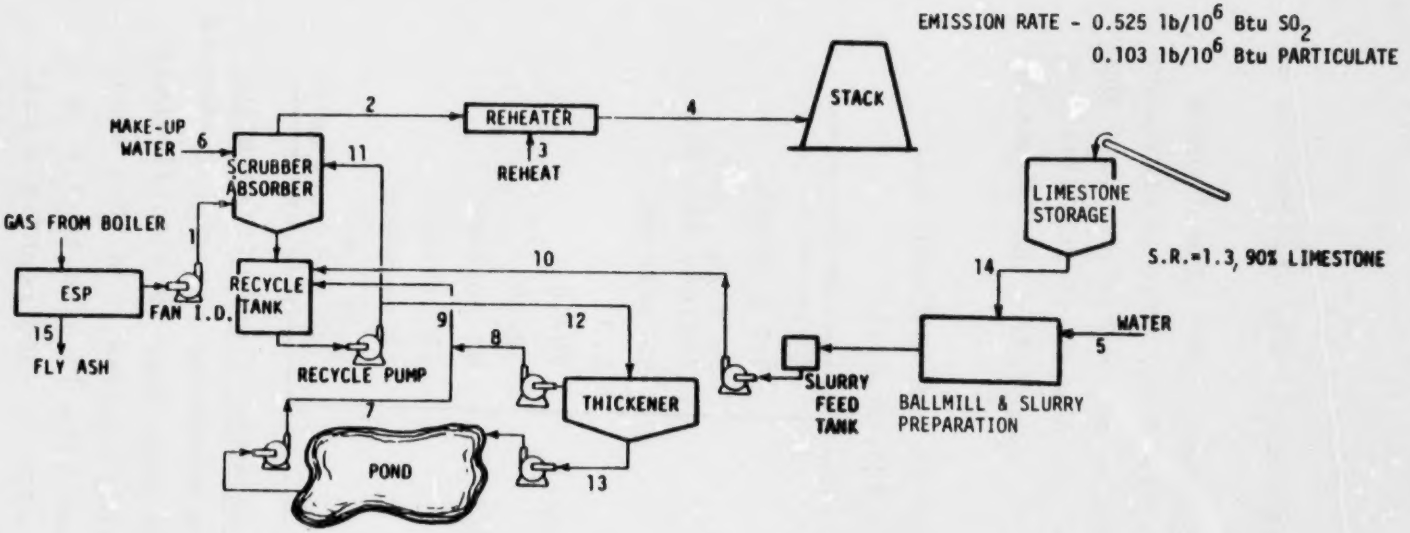
3.2.1.1 Process Flow/Material Balance - The flow diagram for a model plant is shown in Figure 3-26. The assumptions used to make this balance are presented in Table 3-10. This flow diagram is identical to the one presented for lime slurry scrubbing in Section 3.1.1, except for the limestone slurry preparation equipment.

Table 3-10. ASSUMPTIONS USED IN THE MATERIAL BALANCE CALCULATIONS FOR A 500-MW BOILER

SO_2 removal in absorber	90 percent
Coal sulfur content	3.5 percent
Coal heat value	28 kJ/g (12,000 Btu/lb)
Limestone $CaCO_3$ content	90 percent
Stoichiometry	1.3
Sludge oxidation ($CaSO_3$/$CaSO_4$)	90/10
Fly ash/bottom ash	80/20
Particulate removal in ESP	99 percent
Excess air to air heater	40 percent
Excess air to boiler	10 percent
L/G	5.36 l/m^3 (40 gal/1000 scf)

3.2.1.2 Process Chemistry - Limestone Slurry Preparation - The first step in the limestone scrubbing process, limestone slurry preparation, is the only step that is significantly different from the lime scrubbing process. In this step, the limestone is taken from lump size [2.5 cm to 7.6 cm (1 in. to 3 in.)], pulverized to -200 to -300 mesh in a ball mill and slurried with water.[4] It is necessary to grind to

EMISSION RATE - 0.525 lb/10^6 Btu SO_2
0.103 lb/10^6 Btu PARTICULATE



Description-gas streams	1 Gas to scrubber	2 Flue gas to reheat	3 Reheat	4 Gas to stack
Rate, 10^3 lb/hr	5570	5872	1380	7252
SCFM 10^3	1121	1224	299	1523
Temp., °F	320	135	400	175
Fly ash, 10^3 lb/hr	0.5	0.5	0	0.5
SO_2, 10^3 lb/hr	25.5	2.55	0	2.55
H_2O, 10^3 lb/hr	20.1	28.5	12.4	20.8
N_2, 10^3 lb/hr	2970	3970	1050	5020
CO_2, 10^3 lb/hr	1030	1046	0	1046
O_2, 10^3 lb/hr	343	342.4	320	662.4

Description liquid streams	5 Water to Ballmill	6 Makeup water	7 Pond water return	8 Thickener overflow	9 Return water	10 Slurry to recycle	11 Scrubber recycle	12 Spent slurry to the thickener	13 Spent slurry to pond
Rate, 10^3 lb/hr	96.2	217.8	95.7	275	370.7	148	32037	434.5	159.5
GPM	192	435	191	549	740	231	58175	789	260
Solids, %	0	0	0	0.5	0.38	35	15	15	40.0
Specific gravity	1.00	1.00	1.00	1.00	1.00	1.28	1.1	1.1	1.325
Temp., °F	70	70	70	125	70	70	125	125	125
$CaSO_3 \cdot 1/2\ H_2O$ 10^3 lb/hr	0	0	0	0.91	0.91	0	3134	41.6	41.6
$CaSO_4 \cdot 2H_2O$ 10^3 lb/hr	0	0	0	0.14	0.14	0	466	6.2	6.2
H_2O 10^3 lb/hr	96.2	217.8	95.7	273.6	369.3	96.2	27231	369.3	95.7

Description-solid streams	14 Limestone	15 Fly ash
Rate, 10^3 lb/hr	51.8	65.5

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Figure 3-26. Flow diagram/material balance - 500-MW limestone scrubbing system.

this mesh, because limestone slurry is less reactive than lime slurry under similar circumstance. The limestone slurry reaction requires the dissolution of $CaCO_3$; this is a slow reaction. The rate of reaction is directly related to the particle surface area and therefore to the particle size.

SO_2 Reactions

The reactions are the same as the reactions detailed in the lime slurry scrubbing section, except the reactant species is $CaCO_3$, not $Ca(OH)_2$. The reactions are outlined below.[5]

$CaCO_{3\ (s)} \leftrightarrow CaCO_{3\ (aq)} \leftrightarrow Ca^{++} + CO_3^{--}$

$CO_3^{--} + CO_2 + H_2O \leftrightarrow 2HCO_3^-$

$SO_{2\ (g)} \leftrightarrow SO_{2\ (aq)} + H_2O \leftrightarrow H^+ + HSO_3^-$

$HSO_3^- \leftrightarrow H^+ + SO_3^{--}$

$Ca^{++} + SO_3^{--} + 2\ H_2O \leftrightarrow CaSO_3 \cdot 2H_2O_{(s)}$

$H^+ + HCO_3^- \leftrightarrow H_2CO_{3\ (aq)} \leftrightarrow CO_{2\ (g)} + H_2O$

$CaSO_3 \cdot 2H_2O + 1/2\ O_2 \leftrightarrow CaSO_4 \cdot 2H_2O_{(s)}$

$HSO_3^- \leftrightarrow H^+ + SO_3^{--}$

$Ca^{++} + SO_3^{--} + 1/2\ H_2O \leftrightarrow CaSO_3 \cdot 1/2\ H_2O_{(s)}$

$H^+ + HCO_3^- \leftrightarrow H_2CO_{3\ (aq)} \leftrightarrow CO_{2\ (g)} + H_2O$

$CaSO_3 \cdot 2H_2O + 1/2\ O_2 \leftrightarrow CaSO_4 \cdot 2H_2O$

3.2.2 Domestic Units

The first large domestic limestone scrubbing units injected limestone into the boiler and often used a tail-end scrubber to complete SO_2 removal. The first system was the Meremac No. 2 unit of Union Electric Corporation, which started up in 1968; the second was the Lawrence No. 4 unit of Kansas Power & Light, which started up in 1971. The Meremac unit was eventually shut down and the scrubber at Lawrence was converted to a conventional wet scrubbing system.[6]

Recently, all major limestone units have employed tail-end scrubbing. A list of these domestic units appears in Table 3-11. The major facilities are described in Appendix B. It is important to note that each of these limestone scrubbing facilities that has undergone performance testing has met or exceeded design specifications. Design specifications usually call for 60- to 80-percent SO_2 removal efficiency, which is sufficient to meet local codes. Three units, La Cygne No. 1 and Sherburne No. 1 and No. 2, are described in detail in this section.

3.2.2.1 La Cygne No. 1 - Kansas City Power & Light Co. - The LaCygne Power Station is about 88 km (55 miles) south of Kansas City, in Linn County, Kansas. The electric power generating facilities of La Cygne No. 1 consist of a coal-fired, base-loaded boiler with a capacity of 2,800,000 kg steam/hr (6,200,000 lb steam/hr) with an associated 820-MW (net) steam turbine and electric generator. The plant also has three oil-fired boilers, used primarily for start-up of the large unit. The power generating facilities went into service May 31, 1973.

The boiler at La Cygne No. 1 was designed by B&W and is a wet-bottom, cyclone-fired unit. The pollution control equipment consists of eight scrubbing modules, also built by B&W as an integral part of the power generating facility. It is not possible to bypass the boiler's flue gas around the FGD system. The La Cygne Power Station uses about 24 MW from its gross generating capacity of 870 MW to operate the the emission control system. Another 30 MW are required to operate auxiliary station equipment.

Table 3-11. MAJOR DOMESTIC FGD INSTALLATIONS - LIMESTONE SLURRY[7]

Station/Unit Power company	Size (MW)	Start-up date	New or retrofit	Coal percentage sulfur	Design SO_2 removal efficiency, %
Cholla No. 1 Arizona Public Service	115	10/73	Retrofit	0.4-1	92[a]
Duck Creek No. 1A Central Illinois Light	400	8/78[b]	New	2.5-3	75
La Cygne No. 1 Kansas City Power & Light	820	2/73	New	5.0	76
Lawrence No. 4 Kansas Power & Light	125	12/68	Retrofit	0.5	76
Lawrence No. 5 Kansas Power & Light	400	11/71	New	0.5	65
Martin Lake No. 1 Texas Utilities	793	10/77	New	1.0	60
Sherburne No. 1 Northern States Power Co.	710	3/76	New	0.8	50
Sherburne No. 2 Northern States Power Co.	680	9/77	New	0.8	50
Southwest No. 1 Springfield City Utilities	200	4/77	New	3.5	80
Widows Creek No. 8 Tennessee Valley Authority	550	5/77	Retrofit	3.7	80
Will County No. 1 Commonwealth Edison	167	2/72	Retrofit	4.0	82
Winyah No. 2 South Carolina Public Service	280	7/77	New	1.0	70

[a] The A-side train includes the packed absorber with limestone slurry circulated through the module. SO_2 removal for the A-side is 92 percent. The B-side train does not include packing or limestone slurry and removes approximately 25 percent of the SO_2. Total system removal efficiency is 58.5 percent.

[b] One module operated from September 1976 to April 1977.

The coal ranges in gross heating value between 19,000 and 23,800 kJ/kg (8200 and 10,200 Btu/lb). Ash and sulfur range from 20 to 30 percent and 5 to 6 percent respectively.[8] Table 3-12 presents pertinent data on plant design and operation.

Table 3-12. POWER PLANT AND FGD SYSTEM DESIGN/ OPERATING DATA, LA CYGNE NO. 1[9,10]

Maximum generating capacity	820 MW (net)
Boiler manufacturer	B&W
Year placed in service	1973
Maximum coal consumption	366 metric ton/hr (404 ton/hr)
Maximum heat input	8,105 10^6 kJ/hr (7,676 10^6 Btu/hr)
Unit heat rate	9880 kJ/kWh (9,360 Btu/kWh)
Stack height above grade	213 m (700 ft)
Flue gas rate - maximum	1,297 m^3/s (2,760,000 acfm)
Flue gas temperature	141°C (285°F)
Particulate	
Removal efficiency (actual)	97 to 99 percent
SO_2	
Removal efficiency (actual)	70 to 83 percent
No. of FGD modules	8
Process vendor	B&W

Flue Gas Scrubbing System

The FGD system consists of eight identical scrubbing modules (Figure 3-27) with a venturi scrubber for particulate emission control and an absorber tower for SO_2 emission control. Each module treats about one-eighth of the total flue gas from the coal-fired boiler. As the hot flue gas

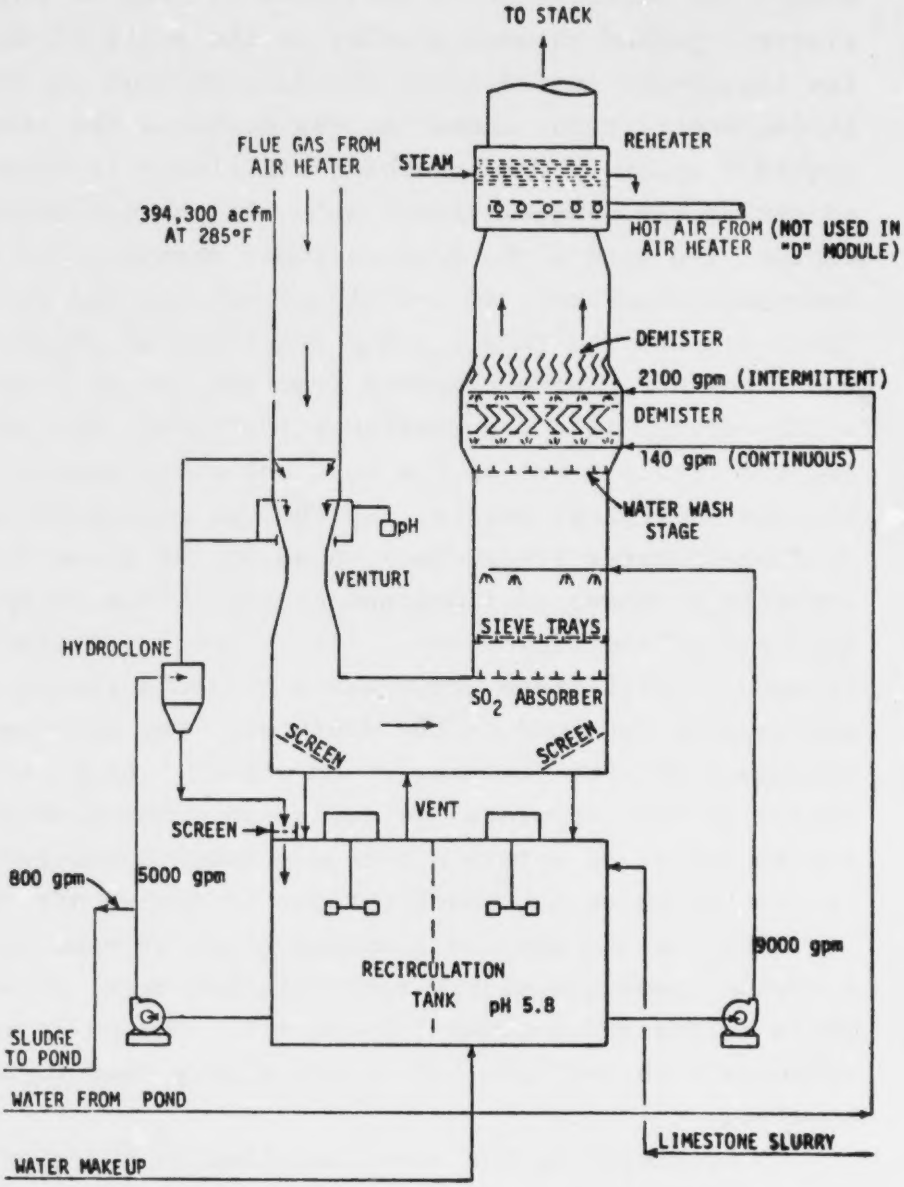

Figure 3-27. Flow diagram of one of the eight FGD modules - La Cygne No. 1.

enters the venturi, it is subjected to jets of limestone slurry injected through nozzles on the walls of the vessel. The liquid-gas stream flows downward through the venturi throat restriction, where the gas contacts the atomized liquid droplets. The scrubbing efficiency is regulated by adjusting the venturi throat gap. As the gas leaves the venturi and enters the disengagement chamber, its velocity decreases from about 40 m/s (130 ft/s) (at the throat) to about 4.6 m/s (15 ft/s). This reduction in velocity separates the limestone slurry droplets from the quenched gas. The slurry drains into the recirculation tank. Gas enters the SO_2 absorption tower at the base and moves upward through two sieve trays in series. As the gas passes through the 34.9-mm-diameter (1-3/8-in.) holes of the sieve trays, it contacts a shower of limestone slurry, which is sprayed into the path of the rising gas. The scrubbed gas then passes through a third sieve tray, which collects slurry carryover and reduces the load on the demister. The gas then passes through a chevron demister 25-cm (10-in.) high, where the remaining fine droplets are collected. The flue gas is then heated before it enters a plenum common to all modules and is discharged to the stack through induced-draft fans.

The venturi and the absorber tower of each module share a common limestone slurry recirculation tank, in which the pH is maintained between 5.5 and 6.0. The pH is monitored by means of a cell located in the slurry feed to the venturi nozzles.

A hydrocyclone has been installed in the slurry circuit of each module to remove large particles, to prevent plugging of nozzles and strainers, and to reduce erosion in pumps, pipes, and nozzles.

The liquid level in the recirculation tank is maintained by pumping excess liquor to the sludge disposal pond. This plant does not use facilities for sludge treatment or fixation.

Limestone is ground on site. A 54,000-metric ton (60,000-ton) supply of limestone rocks is kept near the coal storage area. The limestone is transported intermittently to the mill by the coal conveyor system. Two wet ball mills, each rated at 98 metric tons (108 ton)/hr, are housed in a building that also contains two limestone holding tanks.[11]

Current Design Parameters

The FGD installation was designed with venturi and turbulent contact absorber (TCA) towers for fly ash and SO_2 removal. Many major modifications in the system have since altered some of the original design parameters; the following description represents current operating conditions.

The system was originally designed with seven modules and space was provided for an eighth if needed. Shortly after start-up, it was decided to install the eighth module to allow higher boiler load, and to improve performance and availability.

The venturi scrubber operation was originally designed for an L/G ratio of 2.4 l/m^3 (18 gal/1000 acf) of gas. Since installation of the hydroclones, however, the recirculation pump head has increased so that the L/G is only about 1.6 l/m^3 (12 gal/1000 acf) of gas. The SO_2 absorber tower is designed for an L/G of about 3.5 l/m^3 (26.5 gal/1000 acf).

Gas velocity through the venturi throat is about 150 ft/s with the throat wide open. Gas velocity through the demister section of the tower is 2.6 m/s (8.4 ft/s).

The tower demisters are washed by underspray and overspray manifolds using pond water. Each demister is washed continuously with 8.5 l/s (140 gpm) of underspray water. The overspray operates intermittently at 130 l/s (2100 gpm) for 1 minute during each 8-hour period.

The reheater tube bundles were originally made of 304 stainless steel. The original design reheater exit temperature was 64°C (147°F). Supplemental direct heating with hot air injection had been practiced on six of the original seven modules for gas temperature elevation purposes prior to passage through the reheaters. Additional reheat capacity has been installed in each module to eliminate the need for hot gas bypass.[12]

System Performance

The FGD installation on the La Cygne Boiler No. 1 was plagued with numerous start-up problems. Some of these problems, such as vibrations of the induced-draft fans and their sensitivity to imbalance, appeared even before the boiler was fired and were not due to FGD system operation. An account of the major problems follows.

Scrubbing Modules The process piping network is all rubber-lined to protect the carbon steel base from the abrasive slurry. In general, the rubber liners have performed well, with only a few reported incidents of wear in the area of the spent slurry valves. This wear was attributed to the throttling action of the valve to modulate the flow of slurry. The problem was solved by operating the valve only in a completely open or completely closed position.

The venturi pumps on the circulation tanks are also rubber lined; this lining has been damaged many times, primarily because of plugging of the strainer at the suction end of the pump. As the flow ceases or is drastically reduced because of plugging, the pump cavitates and the liner is sucked into the path of the impeller and shredded. Since the suction strainers plugged frequently, and they were located inside the recirculation tank, the tank had to be drained so the strainers could be cleaned. In order to extend the life of the limestone slurry spray nozzles and to reduce wear and erosion in the slurry recirculation loop, a hydroclone was installed in the recirculation line of each module. Thus it was possible to remove the strainer and correct the pump problems.

In the past, the demister trays and the gas reheater tubes have plugged severely. The demister trays consist of Z-shaped fiberglass boards. Droplets of slurry are carried over with the flue gas and are deposited on the trays. As the slurry builds up, the gas flow is restricted and its velocity through the demister increases. This leads to solids carry-over and deposition on the reheater tubes. The slurry carry-over also reaches the induced-draft fan and is deposited on its blades.

These interrelated problems of carry-over to the demister, the reheater, and the induced-draft fan have necessitated many modifications and corrections of operating procedures. Currently, intermittent heavy overspray and continuous underspray have kept the demisters relatively clean. Steam soot blower modifications have been reasonably successful in the proper maintenance of reheaters, and fan washing has been greatly reduced.

The original reheater tubes (304 SS) began to fail prematurely because of attack by acid condensate. They are being replaced with 316L SS bundles.

Induced-Draft Fans One problem was severe vibration in the fan housing. The fans were initially sensitive to imbalance, and it was found that the operating speed was close to the critical speed. The vibrations caused cracks in the inlet cones, requiring additional stiffeners to strengthen the housing. The thick center plate was scalloped to hold down the weight, and the critical frequency was moved farther away from the operating speed to reduce the tendency to vibrate. The leading edge of each blade was covered with a stainless steel clip to prevent erosion. Particulate carry-over still requires that the fans be washed on an intermittent basis, but the cleaning frequency is being steadily reduced.

The high running temperature of the thrust collars on the fan bearings presented another problem. Temperatures reached the 82°C (180°F) alarm set point within a short time after the fans were started. Temperature increases were controlled by cutting oil grooves in the thrust collar and installing forced-lubrication systems on all the fans. These modifications caused the thrust collar temperature to drop into a range from 60°C (140°F) to 71°C (160°F).

Performance History Despite the problems at start-up, the availability of the system improved steadily. This system is now one of the most reliable large domestic utility FGD systems. Figure 3-28 summarizes availability data. As shown, availability for 1976 averaged 91 percent; the first half of 1977 averaged about 93 percent.[13]

This system was designed for 76-percent SO_2 removal efficiency. Actual SO_2 efficiency has been 80.18 percent with the seven modules operating on 720 MW (Table 3-13). Under maximum load, the removal efficiency averaged 76.2

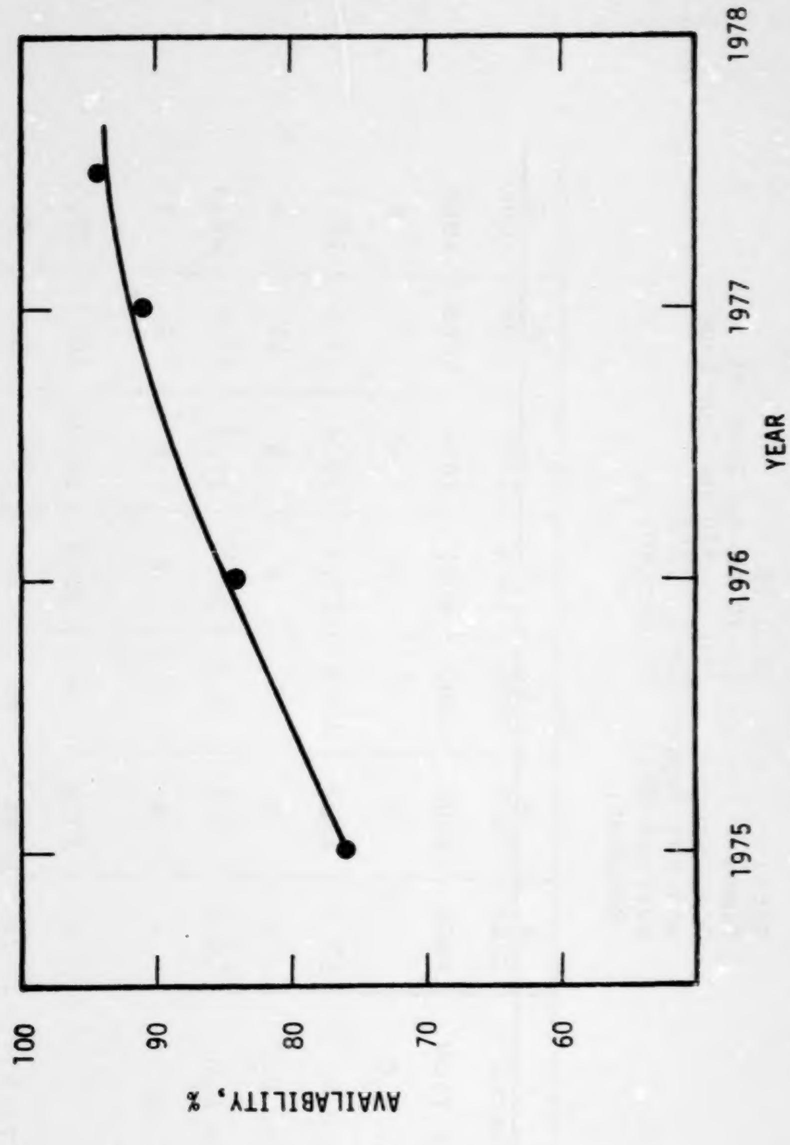

Figure 3-28. La Cygne availability history.

Table 3-13. LA CYGNE STATION STACK SAMPLING TEST - SEVEN MODULES OPERATING[14]

Date: 5/15/75
Time: 11:30 PM to 8:00 PM
Load Range: 700 to 720 MW Continuous
Ambient Temp: 22°C (72°F)
Average SO_2 Removal: 80 percent

	A	B	C	D	E	F	G
Gas flow indicated	320K	280K	300K	340K	320K	300K	300K
Venturi slurry flow	4000	4000	4000	4000	4000	4000	4000
Venturi Δp in. H_2O	7	6	7	7	8	7	8
cm H_2O	17.8	15.2	17.8	17.8	20.3	17.8	20.3
Reheater Δp in. H_2O	4	3	3	8	6	11	8
cm H_2O	10.2	7.6	7.6	20.3	15.2	27.9	20.3
Absorber Dem. Δp in. H_2O	-	4	-	8	8	8	8
cm H_2O	-	10.2	-	20.3	20.3	20.3	20.3
SO_2 efficiency, %	76	81	81	82	77	83	81
Inlet, ppm	4506	4297	4663	4276	4982	4156	834
Outlet, ppm	1068	834	892	776	1121	704	917

percent (Table 3-14). Efficiencies under both conditions should improve now that eight modules are operating.

3.2.2.2 Northern States Power Co. - Sherburne Station No. 1 and No. 2 - The Sherburne County Generating plant of the Northern States Power Company is adjacent to the Mississippi River in Sherburne County, near Becker, Minnesota. Each of the two units has a net capability of 700 MW. The boilers are controlled circulation, single-reheat, balanced draft units manufactured by Combustion Engineering. Two additional units are also being planned to boost the total plant output to 3100 MW by 1983.

The coal presently in use is subbituminous western coal. It has 28-percent moisture, 9-percent ash, 0.8-percent sulfur and a high heating value of 19,000 kJ/kg (8300 Btu/lb).[16]

Flue Gas Scrubbing System

The two Combustion Engineering limestone scrubber systems each use twelve scrubber modules, eleven of which are required for full-load operation. The flow through each of the twelve scrubber modules is approximately 94 m^3/s at 154°C (200,000 acfm at 310°F). The approximate inlet loading is 6.8 g/dry m^3 (3.0 gr/dscf) particulate and 700 ppm SO_2. The approximate outlet loading is 0.09 g/dry m^3 (0.04 gr/dscf) particulate and 350 ppm SO_2, yielding an SO_2 collection efficiency of about 50 percent.

The scrubber system has a rod-type venturi throat, a marble bed absorber, a two-stage fiberglass Chevron shaped demister and a tube type reheater.

Limestone is fed into the Allis-Chalmers ball mill, where it is ground to 200 mesh, slurried to 60-percent solids and stored in an unbaffled tank. The tank is agitated

Table 3-14. LA CYGNE STATION SAMPLING TEST - SEVEN MODULES OPERATING - MAXIMUM BOILER LOAD[15]

Date:	March 24, 1975
Time:	6:10 PM to 10:25 PM
Power:	Min. - 800 MW (e) Gross - Peak - 830 MW (e)
Outside Temp:	-3.3°C (26°F)
SO_2 Removal Average:	76.2%

	A	B	C	D	E	F	G
m^3/s	179	159	188	193	188	183	160
Gas flows (cfm)	380,000	360,000	400,000	410,000+	400,000	390,000	340,000
Venturi pump flow gpm	3000	4000	5000	4000	5000	5000	5000
l/s	183	244	305	244	305	305	305
Venturi Δp in. W.C.	8.0	7.5	-	11.0	10.0	11.0	9.0
cm W.C.	20.3	19.1	-	27.9	25.4	27.9	22.9
Reheater Δp (X2) in. W.C.	4.0	3.5	3.0	3.5	1.0	5.0	12.5
(X2) cm W.C.	10.2	8.9	7.6	8.9	2.54	12.7	31.7
Abs-Dem. Δp in. W.C.	-	5.5	-	8.5	8.5	9.0	9.0
cm W.C.	-	14.0	-	21.6	21.6	22.9	22.9
SO_2 efficiency, %	75.1	79.1	72.2	72.0	N/A	82.9	75.7
Inlet, ppm	5700	5138	5516	4995	5700	5017	5120
Outlet, ppm	1419	1075	1533	1220	N/A	857	1243

with an axial-type agitator that has air sparged through nozzles in the agitator blades. Finally, the slurry is diluted and recirculated through limestone slurry control valves with a conventional recirculation loop. The limestone is a low-magnesium, high-grade material with 95-percent $CaCO_3$ content.[17]

The I.D. fans are not considered part of the scrubber and are controlled by the boiler plant. Each fan is about 12 ft in diameter, powered with a 4,500 kW (6,000 hp) variable speed drive. Four fans are used on each generating unit. Fans are carbon steel and manufactured by Arsene and Co. The fans are capable of overcoming a 109-cm (43-in.) H_2O differential pressure across the scrubber. In operation, the pressure drop is slightly more than 51 cm (20 in.) H_2O.[18]

The marble beds have been removed from two scrubber modules, converting them into spray towers for test purposes. In all units, strainers on the discharge of the Worthington recirculation pumps have been bypassed. Instead, screens have been added to the interior of the absorbers. The screens are provided with water sprays to be used intermittently to backflush the screens. The system is not entirely successful and the nozzles plug because of precipitate formation (and subsequent dislodgement) behind the screens. Also, settling occurs in the suction chamber when the scrubber is not in operation. To avoid this, air spargers are being added for use during shutdowns.[19]

In the scrubber there are 54 spray nozzles (9 headers, 6 nozzles per header). The nozzles occasionally plug, owing to inefficient strainers and to erosion. Metal nozzles failed in a very short time. Ceramic nozzles are failing because of erosion in the tangential cone-forming chamber.[20]

The venturi scrubber was a retrofit after completion of initial design, since particulate removal was inadequate without it. High velocity through the venturi has resulted in a large amount of wear. The walls of the venturi are now lined with a 9.5-mm (3/8 in.), 316L stainless steel wear plate.[21,22]

Reheaters are three-pass in-line type, using hot water at 138°C (280°F) directly from boiler feed-water deaerators; they operate with no major problems. A reheat of 4.4°C (40°F) is added to the flue gas. Steam-operated soot blowers are part of the design, to prevent buildup of solids in the reheater and at the gas/liquid interface in the venturi.[23]

The plant has two thickeners, one for each operating unit. They are Dorr-Oliver units 48.8 m (160 ft) in diameter. Last winter, the rake drive on one of the thickeners stopped as a result of an overload. This situation was not detected for a considerable time and the entire thickener filled with sludge. Both units were then piped to one thickener. Since operations were not hampered with only one thickener operating, they now operate only one at a time. Thickener design was originally for 10-percent inlet solids, 25-percent outlet solids, and zero solids in the overflow. As the equipment is now operated, part of the recycled water is used to sluice down an ash hopper in the boiler economizers, and average thickener inlet is less than 5-percent solids. Thickener bottoms are about 12 percent, and overflow is about 1 percent. No flocculants are used.[24]

Overflow from the thickeners is mixed with recycle water from the ponds in a water makeup tank and then pumped to the scrubbers and to the economizer ash hopper sluice with 120-l/s (2,000-gpm) Worthington pumps. The Worthington recirculation pumps are made of Ni-hard, but they are wearing excessively. Almost all impellers have been changed out at least once. For testing purposes, impellers made of 28-percent chrome and rubber-lined steel are on order.[25]

Piping used for recirculated slurry is rubber-lined except at the headers, where fiberglass is used. The system has many reducers and sharp bends, and wear of the rubber lining at these points is a problem.[26]

Control of pH is manual, with operation monitored by Universal Interlok pH instruments using a flow chamber through which a slipstream from the recirculation pump flows.[27] Scrubber blowdown to the thickener is automatic, using Texas Nuclear sensors and Leeds and Northrup controllers.[28]

Recycle from the pond is controlled by maintaining the makeup tank level. Bubble-tube controllers are used, but are unsatisfactory and variable capacitance sensors may be installed to replace them. In the thickener, pH rises to 8 because of continued reaction with limestone and with quicklime in the fly ash.[29]

Northern States Power and Combustion Engineering are sponsoring a long-term optimization program for the scrubber system.[30]

Performance

Availability for Unit No. 1 averaged 85 percent for the 4 months of operation after start-up. For the past 12 months, availability has been in excess of 90 percent. Unit No. 2 has shown even better start-up performance, with operabilities averaging about 93 percent for the first 7 months. These data are shown in Figure 3-29.[31]

Table 3-15 shows pertinent operating data for the first 8 months of operation of No. 1 unit. The SO_2 removal

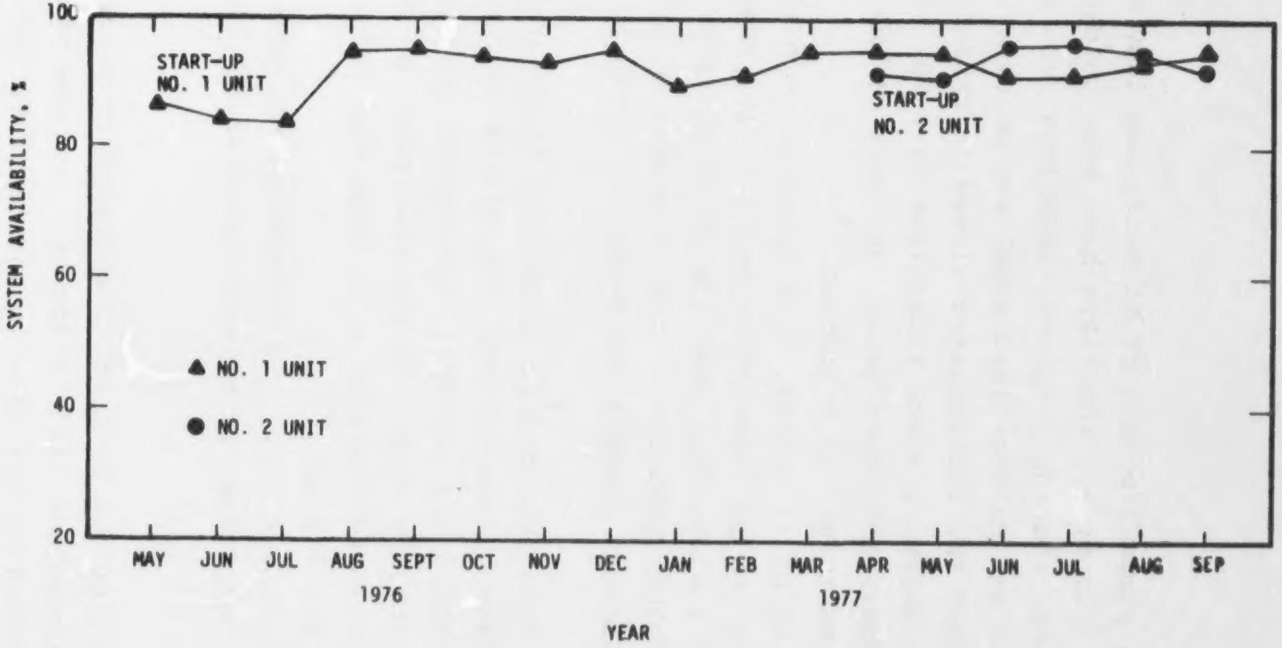

Figure 3-29. Availability history

Sherburne No. 1 and No. 2.

Table 3-15. SHERBURNE COUNTY GENERATING PLANT

UNIT 1 - PERFORMANCE DATA

1. Unit Data (based on May 1 through December 31, 1976 data)	
Electrical output	3,068,130 total MW-hr
Overall capacity factor	60 percent
On-line duration	5,176 hr
II. Scrubber System Data (Averages)	
Particulate concentration:	
Inlet	4.6 to 9.2 g/dry m^3 (2 to 4 gr/dscf)
Inlet	1.7 to 3.4 g/kJ (4 to 8 lb/MM Btu)
Outlet	0.080 to 0.10 g/dry m^3 (0.035 to 0.044 gr/dscf)
Outlet	0.032 to 0.036 g/kJ (0.075 to 0.085 lb/MM Btu)
Removal efficiency	98 to 99 percent
Sulfur dioxide concentration:	
Inlet	400 to 800 ppm
Inlet	0.730 to 0.859 g/kJ (1.7 to 2.0 lb/MM Btu)
Outlet	200 to 400 ppm
Outlet	0.370 to 0.41 g/kJ (0.85 to 0.95 lb/MM Btu)
Removal efficiency	50 to 55 percent

efficiency was 50 to 55 percent, which was sufficient to meet local regulations.[32]

3.2.3 Foreign Units

Table 3-16 lists 17 major limestone slurry scrubbers on boilers in Japan. The Japanese have led FGD technology for several years and it is evident from this table that SO_2 removal efficiencies in the 90- to 95-percent range are common in Japanese limestone scrubbing units. Many of these units are oil-fired and none reports reliability problems. The reasons for the Japanese success are detailed in Section 3.1.3.[33]

3.2.4 Engineering Design Parameters

3.2.4.1 SO_2 Removal Potential of the System - The limitations of the system are identical to the limitations presented for lime scrubbing in Section 3.1 of this report and indicate a theoretical removal efficiency of 100 percent. This section presents the limestone system results as demonstrated at the following actual scrubber installations:

- At the Mohave 170-MW facility of the Southern California Edison Co., SO_2 removal efficiencies in excess of 95 percent were reported for a Turbulent Contact Absorber using limestone slurry for a low-sulfur coal application.[34]

- Reports on the packed-bed module at the 115-MW Cholla facility show 92-percent removal of sulfur dioxide using limestone slurry scrubbing.[35]

- The 10-MW test facility at the TVA Shawnee repeatedly reports SO_2 removal efficiencies in excess of 90 percent. During one run in February 1976, efficiency reached 96 percent in the TCA unit on a high-sulfur coal application.[36]

- The initial operating performance of the Widows Creek No. 8 550-MW scrubber facility indicates that SO_2 removal efficiencies substantially higher than the designed 75 percent value have been measured. These values were measured during manually controlled periods of operation.[37]

Table 3-16. PERFORMANCE OF SO_2 SCRUBBERS USING LIMESTONE SLURRY ON BOILERS IN JAPAN

Plant site	Start-up date	Plant capacity, MW	Fuel	SO_2 concentration, ppm		Removal efficiency, %
				Inlet	Outlet	
Yokosuka	1974	133	Oil	250	40	84
Niigata	1976	140	Oil	550	55	90
Kashima	1976	143	Oil	1000	100	90
Karatsu	1976	243	Oil	550	70	87
Karatsu	1976	190	Oil	550	70	87
Ainoura	1976	243	Oil	880	110	88
Ainoura	1976	243	Oil	880	110	88
Sakata	1976	367	Oil	950	50	95
Sakata	1976	367	Oil	950	50	95
Shimonoseki	1976	400	Oil	1600	50	97
Tamashima	1975	487	Oil	1500	60	96
Omuta	1975	184	Oil	1500	150	90
Takasago	1975	280	Coal	1500	150	90
Takasago	1976	280	Coal	1500	150	90
Isogo	1976	2 x 300	Coal	500	70	86
Fuji	1974	20	NA	1340	20	99
Yokkaichi	1974	83	NA	1600	150	91

- During the St. Clair No. 6 limestone scrubbing demonstration and test program conducted by Detroit Edison and Peabody, SO_2 removal values of 90 and 91 percent (90 percent design) were measured on high (3.0 percent) and low (0.4 percent) sulfur coals.[38]

- At the 167-MW Will County No. 1 facility of the Commonwealth Edison Co., SO_2 removal efficiencies as high as 86.8 percent were measured during a high-sulfur coal (4.0 percent) burn program which yielded inlet SO_2 loadings averaging 3573 ppm.[39]

- SO_2 removal efficiencies in excess of 90 percent have been measured in tests conducted by Combustion Engineering and Kansas Power and Light at Lawrence, Unit No. 4, in 1977. The testing was conducted on the recently installed rod scrubber and spray tower absorber system which replaced the original limestone injection and scrubbing system (1968 start-up). Actual efficiencies in the 95.5 to 97.5 percent range were measured for low-sulfur (0.55 percent) coal.[40]

In summary, current full-scale and test facilities in the United States have demonstrated SO_2 removal efficiencies in excess of 90 percent on both low- and high-sulfur coal. In Japan, results of 95-percent removal from low-sulfur fuel oil are reported.

The ability of an SO_2 scrubbing system to remove high levels of SO_2 is dependent on many design factors. The operating conditions and results at the above plants vary widely, but several key design parameters have proved important to high SO_2 removal efficiency. These parameters have been discussed for the lime slurry system in Section 3.1 of this report. The following key parameters will be outlined in this section for the limestone system:

- L/G ratio - the mass transfer potential of the system is directly related to the ratio of liquid volume to gas volume. SO_2 removal efficiency is greatly increased by increasing L/G.

- pH - SO_2 removal efficiency is directly related to slurry pH.
- Magnesium ion concentration - limestone slurry systems have shown increases in SO_2 removal efficiency when magnesium is added to the slurry.
- Contact surface area - In all mass transfer operations, the rate of transfer is directly related to the transfer area. In a TCA limestone scrubber system, the transfer area (and the mass transfer rate) is directly related to the height of the bed of spheres.

3.2.4.2 L/G Ratio - The L/G ratio is the flow of slurry recirculating through the absorber divided by the flow of flue gas. The affect of L/G on SO_2 removal efficiency is shown in Figure 3-30 for the TCA unit at Shawnee.

The SO_2 inlet (2400 to 2900 ppm) is for a high-sulfur eastern coal; gas velocity is 3.17 m/s (10.4 ft/s); Mg ion concentration is 0 ppm and chloride content varied between 3000 and 7000 ppm. In this example, both the predicted (solid lines) and measured (data points) values indicate the strong dependence of SO_2 removal on the L/G ratio. In a TCA, increased L/G does not significantly increase mass transfer, since the mass contact area is determined largely by the height of spheres, and, in this case, the increased SO_2 removal must be attributed to the better gas/liquid interface conditions.

Figure 3-31 shows a similar effect for a TCA absorber at the 170-MW Mohave test facility, where a 99-percent SO_2 removal efficiency was realized on a low-sulfur (0.38%) western coal.

3.2.4.3 pH - The net effect of slurry pH on SO_2 removal efficiency is illustrated in Figure 3-32 for the TCA at the 10-MW Shawnee test facility. The trend predicted by the Bechtel model in the solid lines and the actual data points

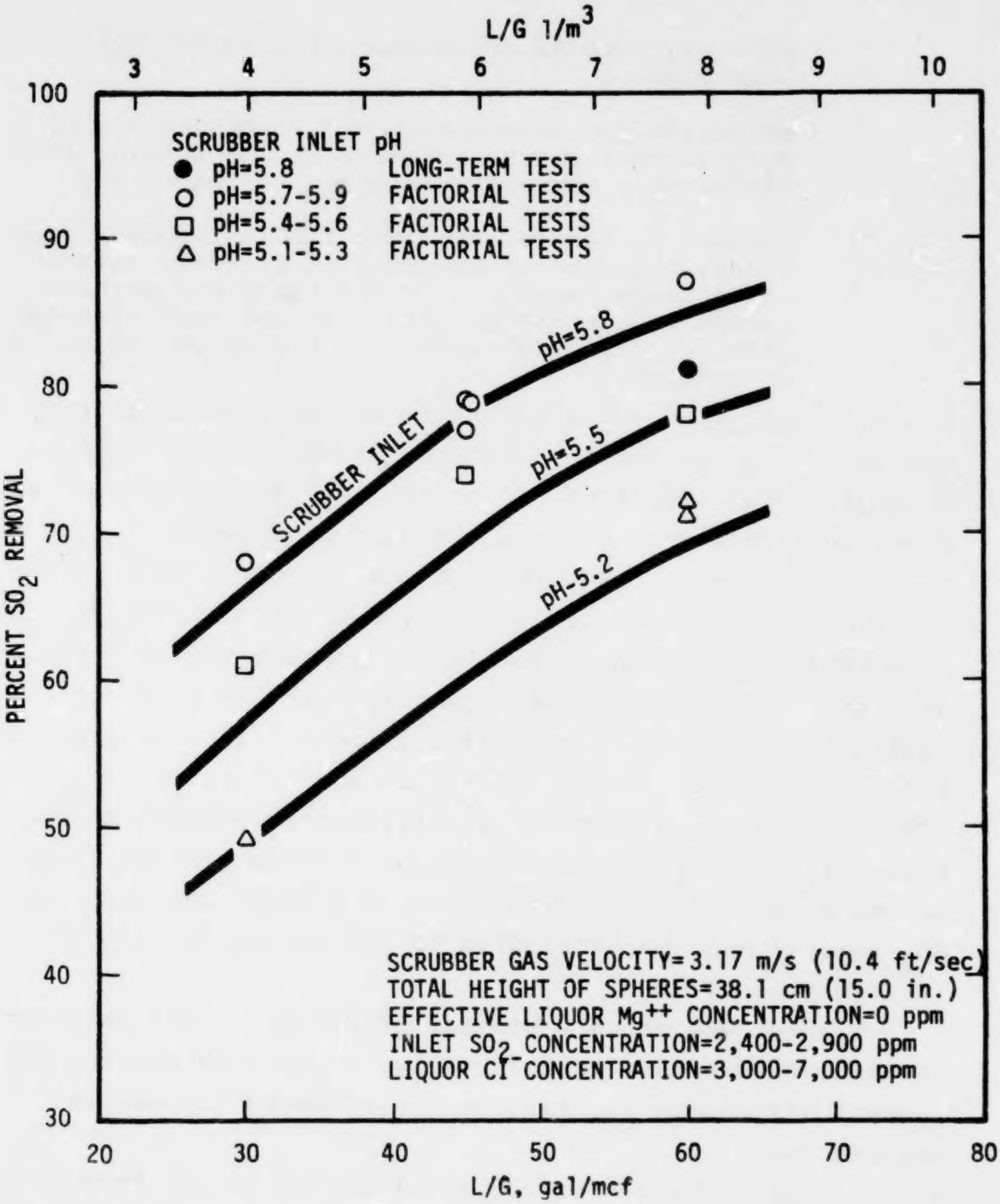

Figure 3-30. L/G ratio and scrubber inlet pH versus predicted and measured SO_2 removal - TCA with limestone - Shawnee plant.[41]

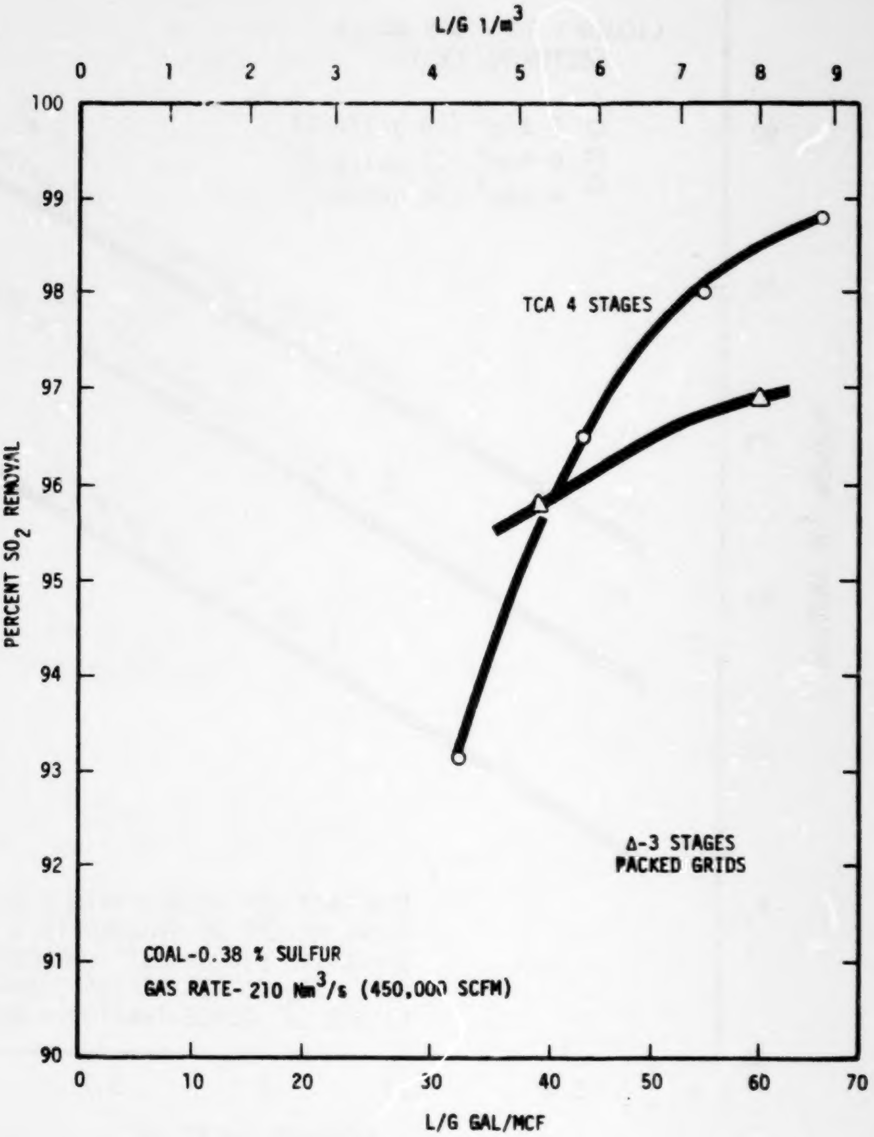

Figure 3-31. L/G ratio versus actual percent SO_2 removal 170-MW test module with limestone - Mohave plant.[42]

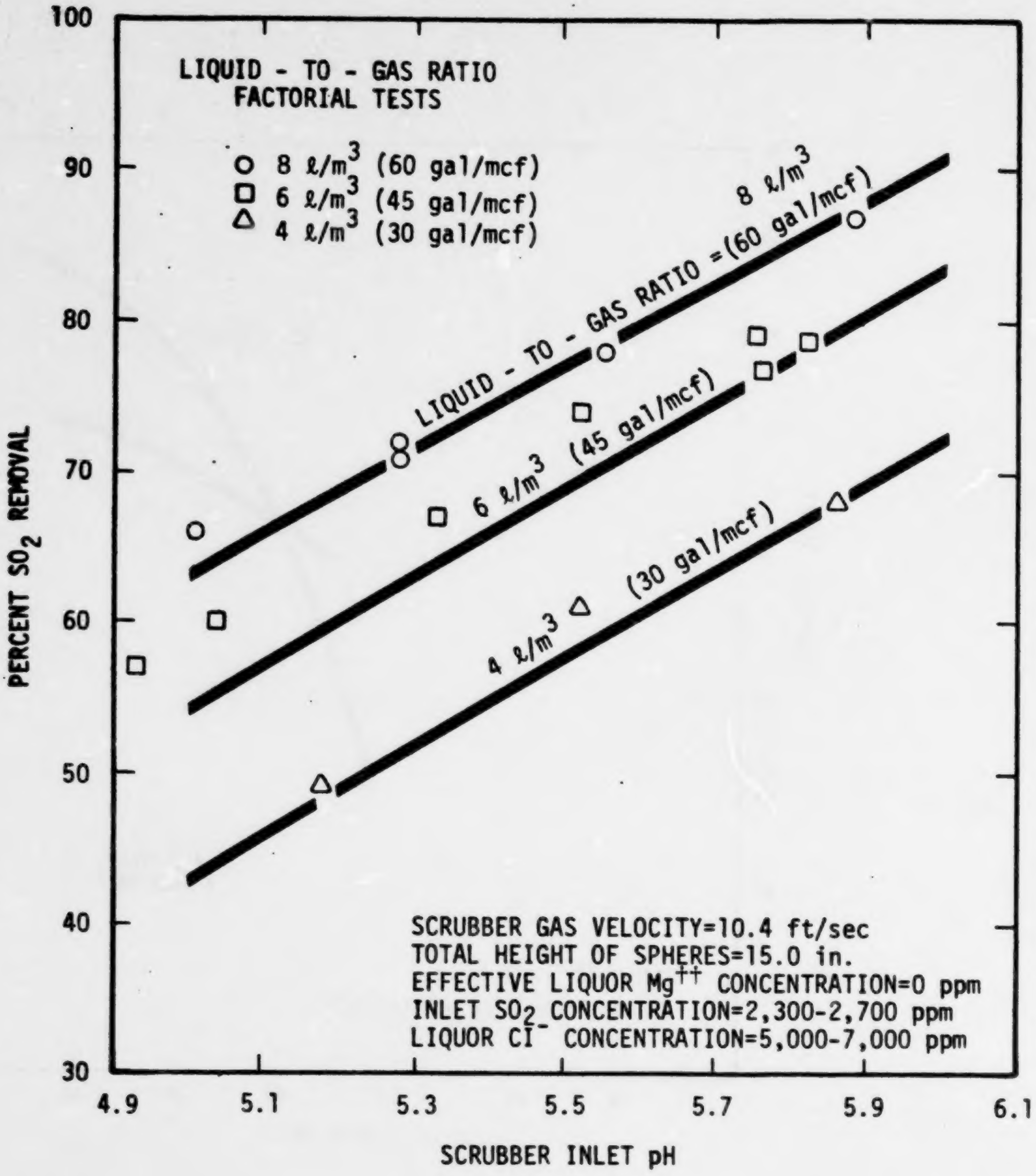

Figure 3-32. Scrubber inlet AV versus PH for three L/G ratios - TCA unit with limestone - Shawnee plant.[43]

are shown. Higher pH results yield higher SO_2 removal efficiencies because of the increased alkalinity of the circulating slurry. Although high pH yields higher SO_2 removal efficiencies, there are some limitations, since high pH also increases scrubber scale formation.

3.2.4.4 Magnesium Ion Concentration - Addition of magnesium ions to the recirculating slurry has demonstrated superior SO_2 removal. The reason for this phenomenon is that the magnesium ionic species are much more soluble than the calcium ionic species. Thus, the magnesium ions are in solution and ready to react with the bisulfite and sulfite ionic species. This high dissolution rate increases the rate at which SO_2 can be removed from the system. The effect of magnesium ion concentration on alkalinity is shown in Figure 3-19.

Actual data demonstrating the effect of magnesium ion concentration on SO_2 removal efficiency are shown in Figures 3-33 and 3-34. These data are for the 10-MW TCA test module at Shawnee. The lower curve on both graphs shows a magnesium concentration of zero. The middle and upper curves show 4700 and 8200 ppm magnesium respectively. It is apparent from these graphs that SO_2 removal efficiencies in excess of 90 percent are possible at moderate [8 l/m^3, (60 gal/1000 acf)] L/G ratios and moderate (5.7) pH values when the recirculating slurry contains magnesium ions.

3.2.4.5 Mass Transfer Area - The rate of mass transfer (and SO_2 removal) is directly related to the mass transfer area. In the limestone slurry TCA system at Shawnee, the transfer area (and SO_2 removal efficiency) is directly related to the height of the bed of spheres (Figure 3-35). Pressure drop also increases with bed height.

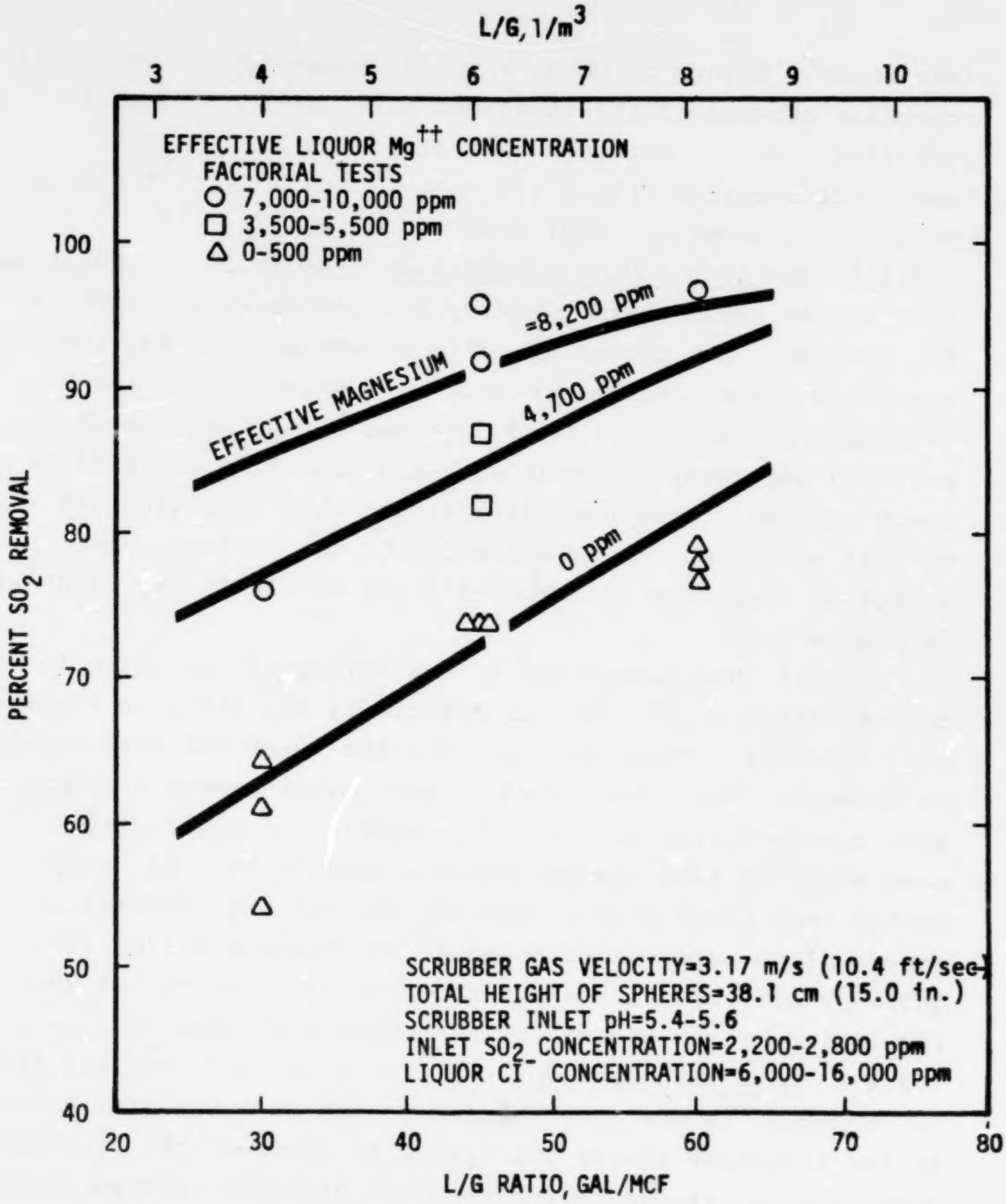

Figure 3-33. L/G ratio versus percent SO_2 removal at various magnesium ion concentrations TCA with limestone - Shawnee plant.[44]

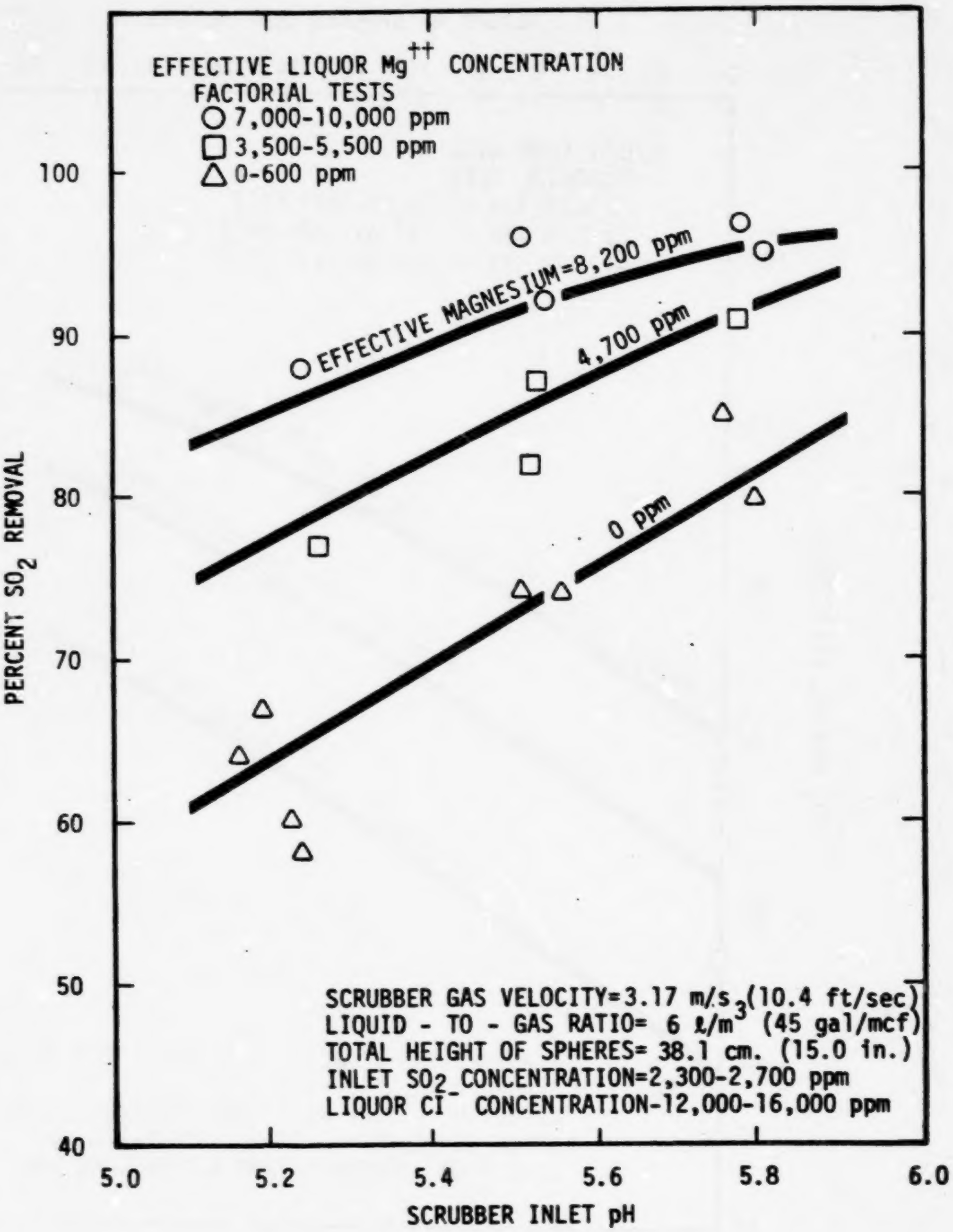

Figure 3-34. Scrubber inlet PH versus percent SO_2 removal at various magnesium ion concentrations TCA with limestone - Shawnee plant.[45]

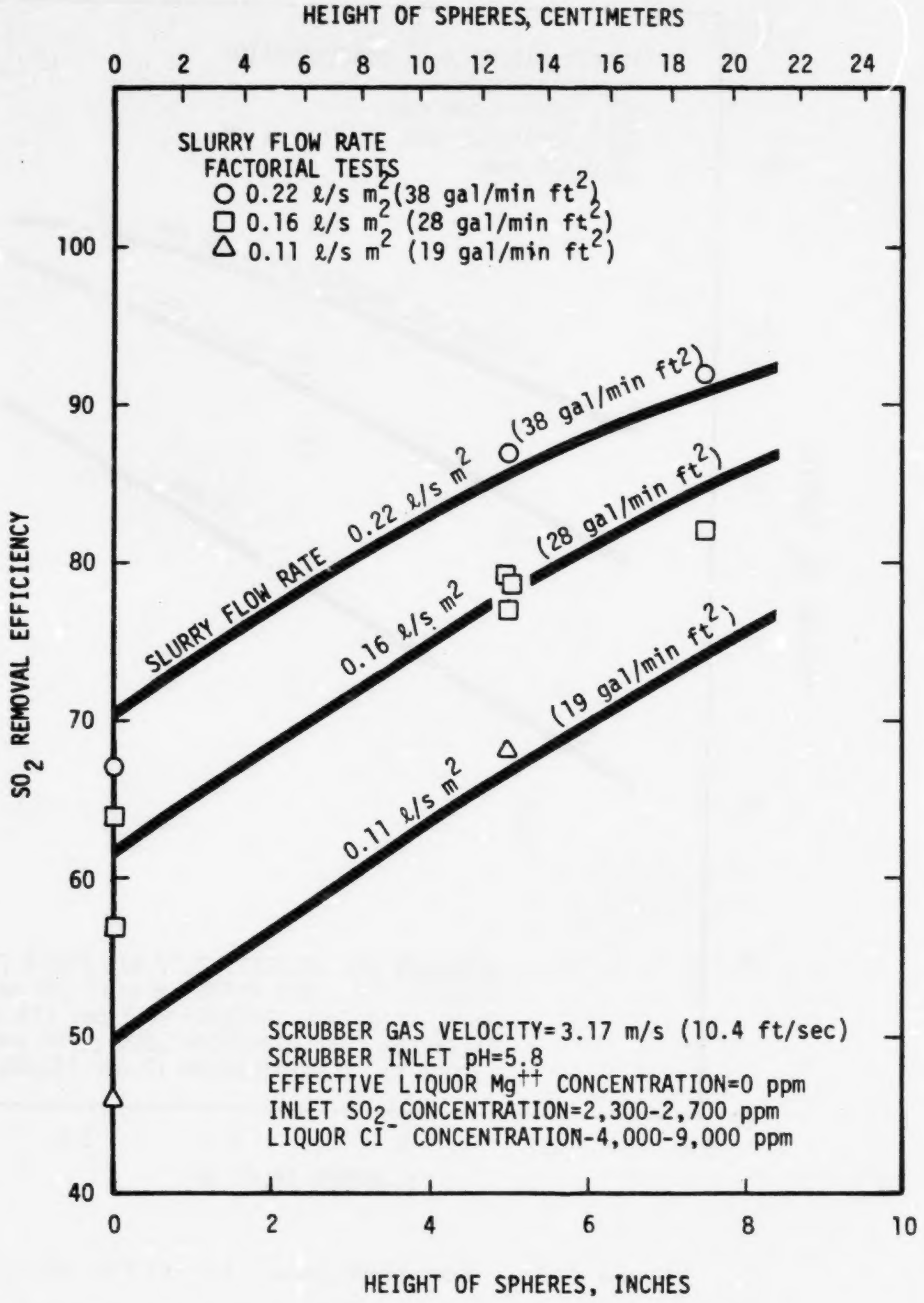

Figure 3-35. Height of spheres versus SO_2 removal efficiency, TCA with limestone - Shawnee facility.[46]

3.2.5 Operability

Although many limestone scrubbing systems have operated poorly to date, several other large-scale commercial limestone scrubbing facilities have demonstrated excellent operability and availability. For example, Figure 3-29 shows availability for the Sherburne No. 1 unit for its 15 months of operation, including start-up. The FGD system has demonstrated monthly system availabilities in excess of 90 percent since shortly after start-up. Also shown is the excellent start-up history for Unit No. 2. Figure 3-28 shows the availability history for La Cygne. This system has steadily improved performance since start-up in 1974 to an average availability in excess of 90 percent for 1976 to 1977.

A combination of good design and maintenance have achieved these high availabilities.

REFERENCES FOR SECTION 3.2

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3.3 DOUBLE ALKALI FLUE GAS DESULFURIZATION SYSTEMS

Double alkali scrubbing is an indirect lime/limestone process that cleans flue gas with fewer of the plugging and scaling problems found in direct lime/limestone systems. The indirect technique involves absorption of SO_2 by an alkaline solution in the scrubbing device, and treatment of the scrubber effluent with lime or limestone in a separate reactor outside the scrubber loop. Calcium sulfites and sulfates are precipitated and discarded, whereas the scrubbing liquor is regenerated and returned to the loop. Figure 3-36 presents a simplified process flow diagram of a typical double alkali scrubbing system.

3.3.1 Process Description

Prior to the removal of sulfur oxides, particulate matter is removed from the flue gas via an upstream ESP, mechanical collector(s), and/or a venturi scrubber. This is done to minimize fly ash erosion of the scrubber internals, and to prevent any appreciable oxidation in the scrubber circuit due to catalytic species in the fly ash.

Sulfur dioxide removal in the scrubber is effected by the reaction of the gas with a circulating alkaline solution sprayed countercurrent to the gas flow. The alkaline reactant employed may be a compound of either potassium, sodium, or ammonia. In the United States, the scrubbing liquor is generally a sodium salt solution.[1]

Soda ash (sodium carbonate) is added to water to form the basic absorbent for the removal of sulfur oxides. Sulfur dioxide and the small amount of SO_3 are absorbed into the scrubbing liquor and react with the sodium carbonate (Na_2CO_3) and sodium hydroxide (NaOH) species to form sodium sulfite (Na_2SO_3), sodium sulfate (Na_2SO_4), and sodium bisulfite ($NaHSO_3$). Some of the SO_2 reacts with sodium sulfite

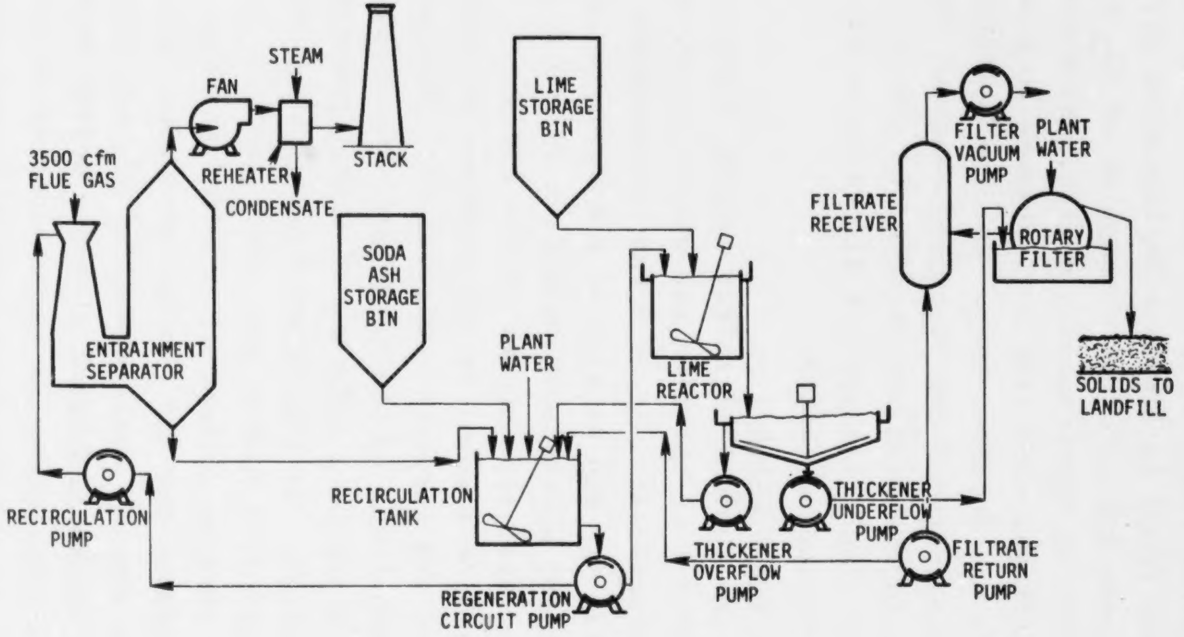

Figure 3-36. FMC double alkali pilot plant: schematic and simplified process flow diagram.

in the presence of water to form sodium bisulfite. An excess of sodium hydroxide, equivalent to about 10 percent of that required to react with all the sulfur oxides in the flue gas, is maintained in the scrubbing liquor. The excess sodium hydroxide reacts with the sodium bisulfite to form more sodium sulfite, which in turn may also be oxidized to form sodium sulfates.

The absorption of sulfur oxides is represented by the following reactions:[2,3]

$Na_2CO_3 + 2SO_2 + H_2O \rightarrow 2NaHSO_3 + CO_2\uparrow$ (1)

$NaOH + SO_2 \rightarrow NaHSO_3$ (2)

$2NaOH + SO_2 \rightarrow Na_2SO_3 + H_2O$ (3)

$2NaOH + SO_3 \rightarrow Na_2SO_4 + H_2O$ (4)

$Na_2SO_3 + SO_2 + H_2O \rightarrow 2\ NaHSO_3$ (5)

$2Na_2SO_3 + O_2 \rightarrow 2Na_2SO_4$ (6)

$4NaHSO_3 + O_2 \rightarrow 2Na_2SO_4 + 2SO_2 + 2H_2O$ (7)

After reaction in the absorbing vessel, spent scrubbing liquor is bled to a reactor tank. Here, inactive sodium bisulfite species are reacted with lime or limestone. In the resultant reactions, sodium bisulfite is converted to sodium sulfite and calcium sulfite, and calcium sulfate solids are precipitated. The reactions vary, depending on whether lime or limestone is used.

Regeneration with limestone (calcium carbonate) takes place over several stages. Sodium bisulfate reacts with the limestone to form sodium bicarbonate ($NaHCO_3$) and calcium sulfite ($CaSO_3$). The solution is acidic because sodium bisulfite is the major reactant. In acidic solution, sodium bicarbonate will break down to sodium hydroxide and CO_2. The sodium hydroxide then reacts with sodium bisulfite to produce sodium sulfite. The series of reactions occurring in the reactor tank when limestone is used as the additive is as follows:[4,5]

$$\mathrm{NaHSO_3 + CaCO_3 + 1/2H_2O \rightarrow NaHCO_3 + CaSO_3 \cdot 1/2H_2O\downarrow} \tag{8}$$

$$\mathrm{NaHCO_3 \rightarrow NaOH + CO_2} \tag{9}$$

$$\mathrm{NaOH + NaHSO_3 \rightarrow Na_2SO_3 + H_2O} \tag{10}$$

When lime (calcium oxide) is used as the regeneration additive, it must first be slaked with water to form calcium hydroxide ($Ca(OH)_2$). Calcium hydroxide reacts with sodium bisulfite to form calcium sulfite and sodium sulfite. Calcium hydroxide also reacts with sodium sulfite and sodium sulfate to form calcium sulfite, calcium sulfate and sodium hydroxide. The series of reactions is as follows:[6,7]

$$\mathrm{NaHSO_3 + Ca(OH)_2 \rightarrow NaOH + CaSO_3 \cdot 1/2H_2O\downarrow + 1/2H_2O} \tag{11}$$

$$\mathrm{Na_2SO_3 + Ca(OH)_2 + 1/2H_2O \rightarrow 2NaOH + CaSO_3 \cdot 1/2H_2O\downarrow} \tag{12}$$

$$\mathrm{Na_2SO_4 + Ca(OH_2) + 2H_2O \rightleftarrows + 2NaOH + CaSO_4 \cdot 2H_2O\downarrow} \tag{13}$$

$$\mathrm{2NaHSO_3 + Ca(OH)_2 \rightarrow Na_2SO_3 + CaSO_3 \cdot 1/2H_2O\downarrow + 3/2H_2O} \tag{14}$$

In regeneration of the scrubbing liquor, hold times in the reactor tank and utilization of limestone and lime depend on solution concentration, temperature, agitation level, and reactant stoichiometry. In general, however, the use of limestone for regeneration requires a hold time of an hour or more to obtain 75 to 85 percent utilization.[8] On the other hand, regeneration with lime requires a hold time of only 10 minutes for utilization of 90 percent or more.[9] The disadvantages of using lime instead of limestone include the necessity of slaking, higher chemical costs, and better pH control requirements.

Following bisulfite removal, a side stream is taken to neutralize the sulfite and sulfate, and to regenerate the original scrubbing liquor. Two distinct methods exist for sulfite/sulfate precipitation.

In most systems, where a dilute concentration ($<0.10M$) of alkali is used in the scrubbing liquor, a lime regeneration method is used. Lime is added to the scrubbing liquor, and reacts with the sodium sulfite and sodium sulfate in solution to form calcium sulfite and calcium sulfate. These reactions are limited by the equilibrium hydroxide ion concentration and the sulfite ion concentration, which is more active than the sulfate species when forming sodium hydroxide. The reactions are as follows:

$$Na_2SO_3 + Ca(OH)_2 + 1/2H_2O \rightarrow 2NaOH + CaSO_3 \cdot 1/2H_2O\downarrow \quad (15)$$

$$Na_2SO_4 + Ca(OH)_2 + 2H_2O \rightarrow 2NaOH + CaSO_4 \cdot 2H_2O\downarrow \quad (16)$$

From the reactor tank, the slurry is pumped either to a clarifier, or to a thickener, where the precipitated solids are separated from the scrubbing liquor. The solids are then delivered to a vacuum filter for further dewatering. From here the sludge may be sent to a chemical fixation tank and eventually to the disposal pond; if fixation is deemed unnecessary, however, they may be disposed of directly.

The second method of sulfate/sulfite removal may be used only in systems employing a concentrated active alkali liquor (>0.15 M). In this process, a side stream from the scrubber discharge is treated with sulfuric acid to precipitate sulfates and regenerate the scrubbing liquor. At present, this regenerating method has only been used in Japan.

For either regeneration method, regenerated liquor is returned to the scrubbing loop after two additional steps. First, makeup sodium must be added to the solution. This is necessary because a small amount of the scrubbing liquor is lost in the waste solids. Second, any excess calcium ion concentration must be reduced in order to prevent scaling in the scrubber. Both calcium sulfate and calcium sulfite are produced during regeneration, as shown in reactions 15 and 16. Since the solubility product of calcium sulfate is approximately two orders of magnitude greater than that of calcium sulfite, calcium sulfate is more likely to remain in solution and to be carried in the liquor from the clarifier. When a gypsum phase is present in the solids at equilibrium (saturated solution), the regeneration system produces high calcium ion concentrations. Therefore, if conditions develop that cause the scrubbing liquor to become supersaturated with calcium sulfate, a high potential for scaling in the scrubber will result.

Any chemical reaction which removes excess minerals from solution in order to reduce scaling potential is known as a "softening" reaction. There are several methods of softening the reactant solution before it returns to the absorber. One of these, known as "sulfite softening", is effected by returning the regenerated liquor to the bisulfite reaction vessel. The high calcium ion concentration is reduced as the calcium combines with residual sulfite in the reactor to form the highly insoluble $CaSO_3 \cdot 1/2H_2O$, which is removed. Sodium carbonate or sodium hydroxide may be added to the liquor to replace lost sodium.

The second method of reducing calcium ion concentrations is known as "carbonate softening". This process involves the use of CO_2 and/or Na_2CO_3.

If limestone regeneration is used, lost sodium is replaced by adding sodium hydroxide or sodium carbonate to the solution; however, only as much sodium hydroxide or sodium carbonate is added to the liquor as sodium has been lost in the waste discharge. The excess calcium might be removed by contacting the solution with a stream of cleaned boiler flue gas containing carbon dioxide, although this has not yet been demonstrated.

The carbon dioxide reacts with the calcium ions as shown in the following equation.

$$Ca^{++} + CO_2 + H_2O \rightarrow CaCO_3 + 2H^+ \qquad (17)$$

The calcium ion concentration is now reduced to a level below the saturation value for $CaSO_4$, and the liquor can be returned safely to the scrubbing loop. The precipitated calcium carbonate settles out for ultimate disposal.

The same procedure for replenishing lost sodium and reducing calcium ion concentrations in the liquor may be used in a lime regeneration system. When lime is used, it is better to add sodium carbonate to the liquor from the clarifier; this not only replaces lost sodium, but also "softens" the solution in the following single chemical reaction.

$$Na_2CO_3 + Ca^{++} \rightarrow CaCO_3\downarrow + 2Na^+ \qquad (18)$$

Sodium is replaced in the scrubbing liquor by the additional sodium hydroxide, and excess calcium is removed as precipitated calcium carbonate. As in the limestone system, ideally only as much sodium is added to the liquor as has been lost in the waste discharge. The amount of sodium carbonate added, however, may be insufficient to soften the solution satisfactorily, and an excess over stoichiometric requirement may be added.

The various alternative steps described above can be combined in a number of ways in double alkali FGD systems. Three basic modes of operation exist for sodium/calcium scrubbing, as defined below:[10]

1. Lime regeneration, concentrated active alkali with side stream sulfate treatment (removal).

2. Limestone regeneration, concentrated active alkali with side stream sulfate treatment (removal).

3. Lime regeneration, dilute active alkali, with carbonate softening.

These modes, their advantages and disadvantages, will be discussed very briefly in the following sections. Many variations on these three basic schemes can be made by changing certain parameters associated with the systems.

1. Lime Regeneration, Concentrated Active Alkali, and Side Stream Sulfate Treatment[11,12]

The first mode of operation involves lime regeneration, use of concentrated alkali solutions, and side stream treatment for sulfate removal and control. Lime is used for the bisulfite neutralization of the concentrated (>0.10 M) alkali scrubbing solutions. A side stream is then taken from the bisulfite reactor, where sulfuric acid is used to precipitate sulfates and regenerate the scrubbing liquor. Sulfite softening is again employed to remove residual calcium ions (Figure 3-37).

This flow scheme has the disadvantage of being prone to sulfate regeneration complications. Its advantages, however, include relatively short hold times necessary for the bisulfite/sulfite regeneration with lime, as well as high reactant utilizations; an advanced state of development; and high SO_2 removal efficiencies with low flows of the concentrated alkali scrubbing liquors. This mode of operation appears to be promising, but its successful application on a universal scale depends on development of an effective treatment for sulfate removal and control.

2. Limestone Regeneration, Concentrated Active Alkali, and Side Stream Sulfate Removal[13,14]

This mode differs from the first only in its use of limestone instead of lime for the bisulfite removal step. Like the previous mode, it too is prone to sulfate regeneration complications. It has the further disadvantage of requiring long hold times for bisulfite/sulfite regeneration. Its advantages, however, appear to be high SO_2 removal

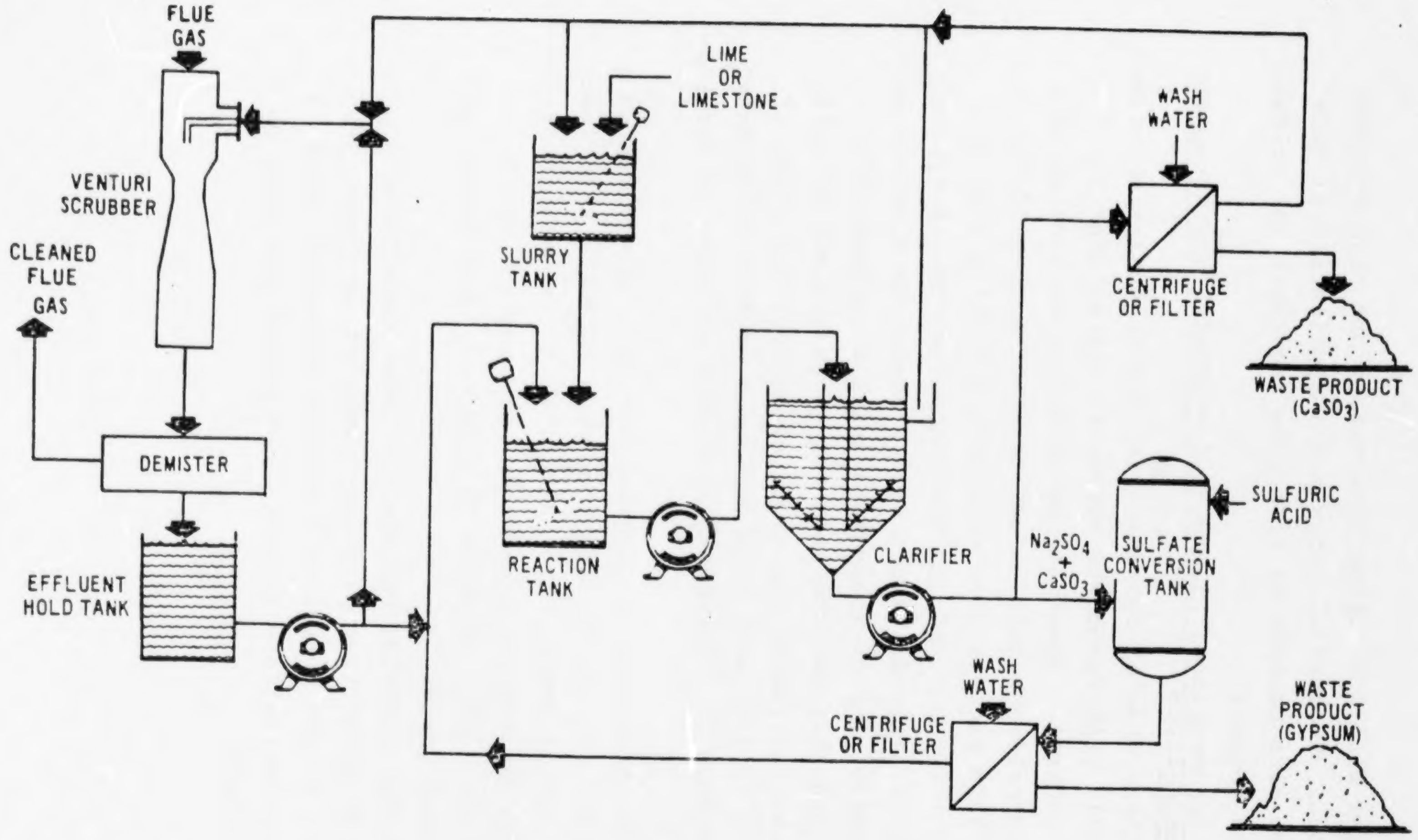

Figure 3-37. Schematic of a double alkali system with lime or limestone regeneration, concentrated alkali, and H_2SO_4 sulfate removal.

efficiencies with low liquor flow rates, and high reactant utilizations. Successful large-scale application of this flow scheme also depends on effective treatment for sulfate removal and control.

3. Lime Regeneration, Dilute Active Alkali, and Carbonate Softening[15,16]

The third mode of operation is lime regeneration using dilute alkali with carbonate softening for calcium ion (scaling) control. Spent scrubbing liquor is treated with lime to neutralize bisulfite and to react with sulfite and sulfate. This process uses carbonate softening (CO_2 and Na_2CO_3) to lower the calcium ion concentration. A typical flow sheet is given in Figure 3-38. Among its disadvantages are the possibility of inadequate sulfate regeneration; scaling potential; and the need for large volumes of relatively dilute scrubbing liquors. Advantages are found in its ability to deal with high oxidation levels, relatively simple equipment requirements, and advanced state of development.

Of the three process schemes described above, the most promising for utility applications appears to be the first, involving use of concentrated alkali with lime regeneration and side stream sulfate treatment. The advantage lies mainly in its higher relative SO_2 removal capabilities and lower capital costs.[17]

A typical material balance for this operational scheme for a 500-MW coal-fired plant is presented in Figure 3-39.

In all of the operational schemes presented above, a calcium sulfite/sulfate sludge is generated that must be disposed safely.

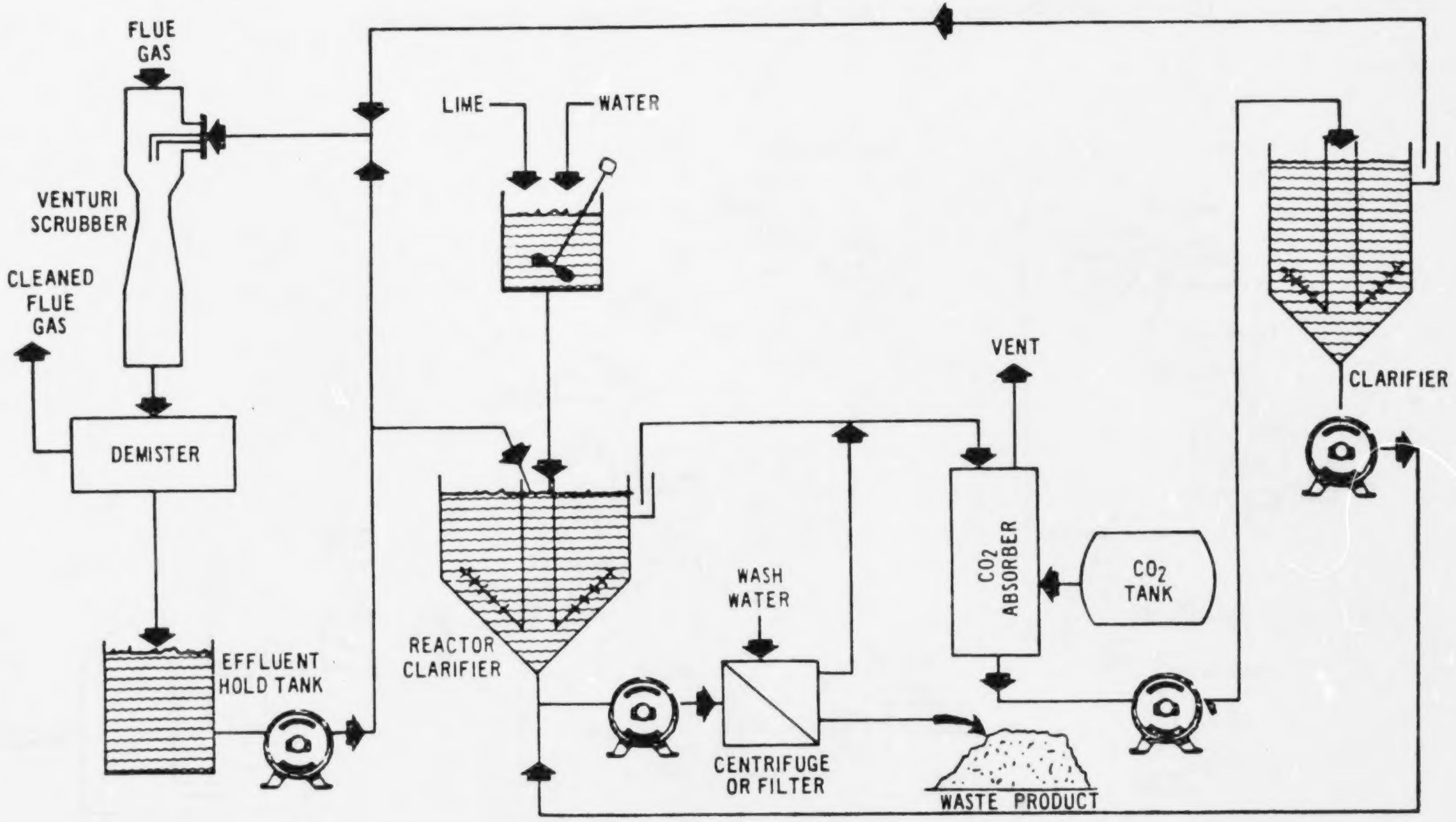

Figure 3-38. Schematic of a double alkali system with lime regeneration, dilute alkali, and carbonate softening.

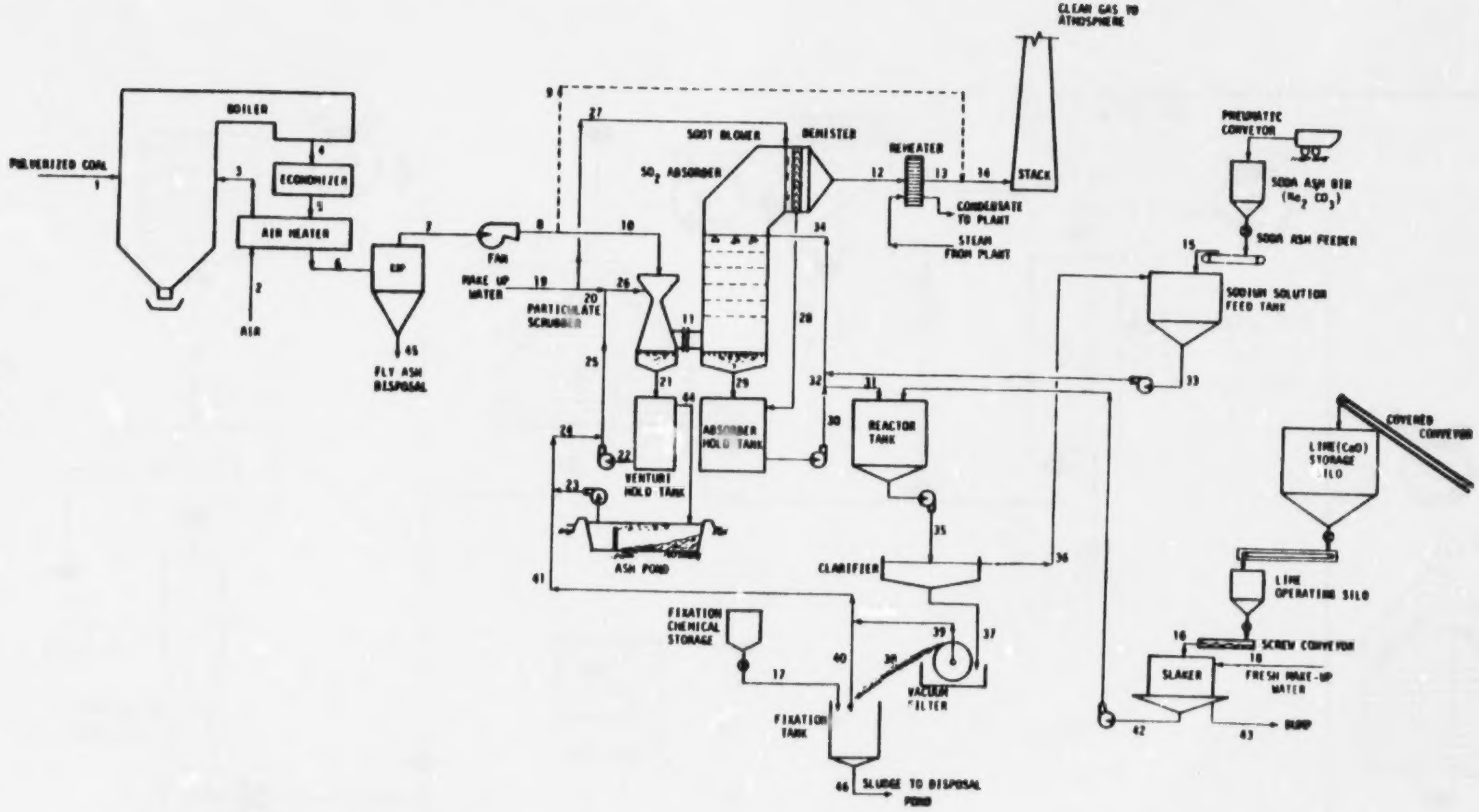

Figure 3-39. Material balance for 500 MW double alkali FGD system with lime regeneration.

Stream no. Description	1 Coal to boiler	2 Air to air heater	3 Air to boiler	4 Gas to economizer	5 Gas to air heater	6 Gas to ESP	7 Gas to fan	8 Fan outlet gas	9 Gas bypass	10 Gas to part. scrbr.	11 Gas to SO_2 ab-sorber	12 Gas to reheater
Rate, 1000 kg/h	170	2049	1848	1998	1998	2199	2186	2186	77.3	2109	2196.6	2180.8
Nm^3/s		464	419	445	445	490	490	490	17	473	503	501.4
m^3/s, Liquid												
Particulates, kg/h				15300	15300	15300	2600	2600	91	2510	12.2	12.2
Temperature, °C		43.3	279	477	374	154					51.7	51.7
Specific gravity												
Viscosity, pascal-second												
Undissolved solids, %												
pH												
Sulfur as SO_2, kg/h	11900			11900	11900	11900	11900	11900	440	11480	10330	770

Stream no. Description	13 Reheater discharge gas	14 Gas to stack	15 Soda ash to feed tank	16 Lime to slaker	17 Chemical to fixation tank	18 Fresh makeup water	19 Fresh makeup water	20 Makeup water to venturi	21 Part. scrbr. slurry to hold tank	22 Recycle from hold tank
Rate, 1000 kg/h	2180.8	2258.1	0.790	11.46	4.58	62	50.76	44.16	4515	4465
Nm^3/s	501.4	518.4								
m^3/s, Liquid						0.0172	0.041	0.0123	1.218	1.204
Particulates, kg/h	103.2	103.2							225750	223250
Temperature, °C	78.9	78.9								
Specific gravity						1.0	1.0	1.0	1.03	1.03
Viscosity, pascal-second										
Undissolved solids, %						-0-	-0-	-0-	5	5
pH										
Sulfur as SO_2, kg/h	770	1190							103700	102520

Figure 3-39 (continued). Material balance for 500 MW double alkali FGD system with lime regeneration.

Stream no.	23	24	25	26	27	28	29	30	31	32
Description	Pond recycle water			Scrubber to part. scrbr.	Demister wash water	Demister wash to hold tank	Spent slurry to hold tank	Recycle from hold tank	Purge to reactor tank	
Rate, 1000 kg/h	45.44	88.24	4553.24	4597.6	6.6	6.6	4900.2	4906.8	110.2	4796.6
Nm^3/s										
m^3/s, Liquid	0.0126	0.0245	1.228	1.240	0.00183	0.00183	1.361	1.363	0.0306	1.332
Particulates, kg/h			223250	223250						
Temperature, °C										
Specific gravity	1.0	1.0	1.03	1.03	1.0	1.0		1.0	1.0	1.0
Viscosity, pascal-second										
Undissolved solids, %	-0-	-0-	5	5		-0-				
pH										
Sulfur as SO_2, kg/h			102520	102520						

Stream no.	33	34	35	36	37	38	39	40	41	42
Description	Soda solution	Scrubbing slurry to absorber	Clarifier feed	Clarifier overflow	Clarifier underflow	Filter cake	Filtrate	Mixing water		Slurry to reactor
Rate, 1000 kg/h	92.4	4889	183.3	91.6	91.7	45.8	45.9	3120	42.8	73.13
Nm^3/s										
m^3/s, Liquid	0.0257	1.3581	0.0467	0.0254	0.0212	0.00848	0.01275	0.00087	0.0119	0.0185
Particulates, kg/h										
Temperature, °C										
Specific gravity	1.0	1.0	1.09	1.0	1.20	1.5	1.0	1.0	1.0	1.1
Viscosity, pascal-second										
Undissolved solids, %	-0-		15	-0-	30	60	-0-	-0-	-0-	20
pH	12.3									12.1
Sulfur as SO_2, kg/h										

Stream no.	43	44	45	46
Description	Slaker dump	Particulate slurry product	Fly ash disposal	Sludge to disposal pond
Rate, 1000 kg/h	0.340	50	12.7	53.5
Nm^3/s				
m^3/s, Liquid		0.0135		0.0099
Particulates, kg/h		2500	12700	
Temperature, °C				
Specific gravity		0.03		1.5
Viscosity, pascal-second				
Undissolved solids, %				60
pH				
Sulfur as SO_2, kg/h				

Notes:

Calculations based on:

a. 130% stoichiometric lime: 95% CaO and 5% grit.
b. 60% of the grit is removed.
c. 90% SO_2 removal.
d. 90% solid in ash pond.
e. $CaSO_3 \cdot 1/2\ H_2O/CaSO_4 \cdot 2H_2O$ = 40/60 by weight.

Figure 3-39 (continued). Material balance for 500 MW double alkali FGD system with lime regeneration.

3.3.2 Double Alkali Scrubber Units in the United States

Several successful bench-scale, pilot plant and prototype double alkali FGD systems have been tested on boiler flue gas applications in the United States. The success of these programs has resulted in commitments by three separate electric utility companies to install full-scale double alkali FGD systems on coal-fired boilers. As yet no full-scale system is operating on a utility boiler in the United States; but several systems are working on coal-fired industrial boilers and one pilot plant system and one prototype have been tested on utility units.

Industrial applications on coal-fired boilers are as follows:

Company: General Motors, Inc.
Plant: Chevrolet
Location: Parma, Ohio
Stream treated: Off-gas from coal-fired boilers
System size: 124 m^3/s (262,000 acfm) (32 MW)
SO_2 inlet: 800 to 1300 ppm (1.5 to 3.0% S coal)
Start-up date: March 1974

Company: Caterpillar Tractor Co.
Plant: Joliet Plant
Location: Joliet, Illinois
Stream treated: Off-gas from coal-fired boilers
System size: 48.8 m^3/s (103,500 acfm) (18 MW)
SO_2 inlet: 2300 ppm (4% sulfur coal)
Start-up date: September 1974

Company: Firestone Tire and Rubber Co.
Plant: Pottstown Plant
Location: Pottstown, Pennsylvania
Stream treated: Off-gas from a power boiler
System size: 6.6 m^3/s (14,000 acfm)
SO_2 inlet: 1,000
Start-up date: January 1975

Company: Caterpillar Tractor Co.
Plant: Mossville Plant
Location: Mossville, Illinois
Stream treated: Off-gas from 4 coal-fired boilers
System size: 113 m^3/s (240,000 acfm) (57 MW)
Fuel properties: Coal, 3.2 percent sulfur average
Start-up date: October 1975

One prototype and one pilot plant double alkali system have been operated on utility coal-fired boilers:

Utility: Utah Power and Light Co.
Unit: Gadsby Station, Unit No. 3
Location: Gadsby, Utah
Unit size: 1.2 m^3/s (2500 acfm) (≈0.6 MW)
Fuel properties: Coal, 0.4 percent sulfur average
Start-up date: 1971
Note: Terminated 1973

Utility: Gulf Power Co.
Unit: Scholz, Unit No. 1
Location: Chattahoochee, Florida
Unit size: 35 m^3/s (75,000 acfm) (20 MW)
Fuel properties: Coal, 3 to 5 percent sulfur
Start-up date: February, 1975
Note: Terminated July 1976

As a result of the success of pilot and prototype systems, three full-scale double alkali systems are scheduled for operation soon on new coal-fired utility boilers. Tables 3-17, 3-18, and 3-19 present design information on these units.

The following sections present detailed information on each of the industrial and prototype units listed above.

3.3.2.1 General Motors - Parma Plant

System Description[22] - Figure 3-40 is a process flow diagram for Parma. Flue gases from four spreader stoker boilers rated at a total of 32 MW are routed to four tray-type absorbers. The gases enter through a prequench-section at the bottom of each scrubber, and then flow in counter-current to an aqueous sodium hydroxide sulfite/bisulfite solution. The scrubber is a four-tray, impingement-type

Table 3-17. DESIGN PARAMETERS FOR THE CANE RUN NO. 6 DOUBLE ALKALI SCRUBBER PLANT,[18] LOUISVILLE GAS & ELECTRIC CO., LOUISVILLE, KY

FGD unit rating, MW	277
Fuel, characteristics	Coal: 26,700 J/g (11,500 Btu/lb) 3.5 to 4.0 percent sulfur, 11.5 percent ash
FGD system supplier	CEA/ADL
Process	Double alkali, sodium-calcium
Active alkali mode	Concentrated
Application	New
Start-up date	2/79
FGD modules	Two
Module design	Two-stage tray tower absorber
System performance guarantees:	
SO_2 removal efficiency	200 ppm in the scrubber discharge of 95 percent SO_2 removal if the sulfur content of the coal is 5 percent or greater
Particulate emissions	No additional loading over inlet concentration
Sodium consumption	0.045 moles of soda ash makeup per mole of SO_2 removed when maximum coal chloride level is 0.06 percent. 0.05 moles of soda ash additional for each mole of chloride in the coal above the 0.06-percent level
Lime consumption	Maximum 1.05 moles/mole SO_2 removed
Power consumption	1.1 percent of peak operating rate (300 MW)
Filter cake properties	55 percent insoluble solids
System availability	90 percent for a one-year operating period

Table 3-18. DESIGN PARAMETERS FOR THE A. B. BROWN NO. 1 DOUBLE ALKALI SCRUBBER PLANT,[19] SOUTHERN INDIANA GAS & ELECTRIC CO., WEST FRANKLIN, INDIANA

FGD unit rating, MW	250
Fuel, characteristics	Coal: 26,700 J/g (11,500 Btu/lb) 3.53 percent S, 8.76 percent ash, 11.35 percent H_2O
FGD system supplier	FMC
Process	Double alkali, sodium-calcium
Application	New
Start-up date	4/79
FGD modules	Two
Module design	Two-stage disc contractor, cascading liquor absorber
SO_2 removal efficiency, (% design)	85[a]
System performance guarantees:	
Availability	95 percent for 60-day consecutive test run
SO_2 removal	85 percent for 4.5 percent sulfur coal
Chemical consumption	Yes (soda ash and lime)[b]
Power consumption	Yes[b]
Mechanical components	One year for mechanical failures
Sludge disposal	Hauled to off site landfill
Projected unit capacity factor	90 percent for the first year of operation; 70-percent average for 35-year life span

[a] Based on 4.5 percent maximum sulfur content in coal.

[b] Specifics of guarantees not known.

Table 3-19. DESIGN PARAMETERS FOR THE NEWTON NO. 1 DOUBLE ALKALI SCRUBBER PLANT,[20,21] CENTRAL ILLINOIS PUBLIC SERVICE CO., NEWTON, ILLINOIS

FGD unit rating, MW	575
Fuel, characteristics	Coal: 25,300 J/g (10,900 Btu/lb) 4-percent sulfur, 0.2-percent chloride
FGD system supplier	Buell-Envirotech
Process	Double alkali: sodium-calcium
Active alkali mode	Concentrated
Application	New
Start-up date	11/79
FGD system design:	
Modules	Four
Module design	High velocity cocurrent spray tower, countercurrent two-tray-stage mobile ball bed absorber (series arrangement)
Regeneration	Three process-causticizers and and two thickeners
Dewatering	Three horizontal extraction filters
Filter cake disposal	Fly ash stabilization/on-site landfill
SO_2 removal efficiency	95 percent, less than 200 ppm
HCl removal efficiency	90 percent
Availability	100 percent for 70-percent load factor for 30-year life span
Soda ash consumption Gg/yr. (TPY)	3.4 (3800) ($\simeq$ 0.05 moles/mole SO_2 removed)
Quicklime Gg/yr. (TPY)	104 (115,000) ($\simeq$ 1.10 moles/mole SO_2 removed)
Power consumption	3 percent of peak operating rate (550 MW)

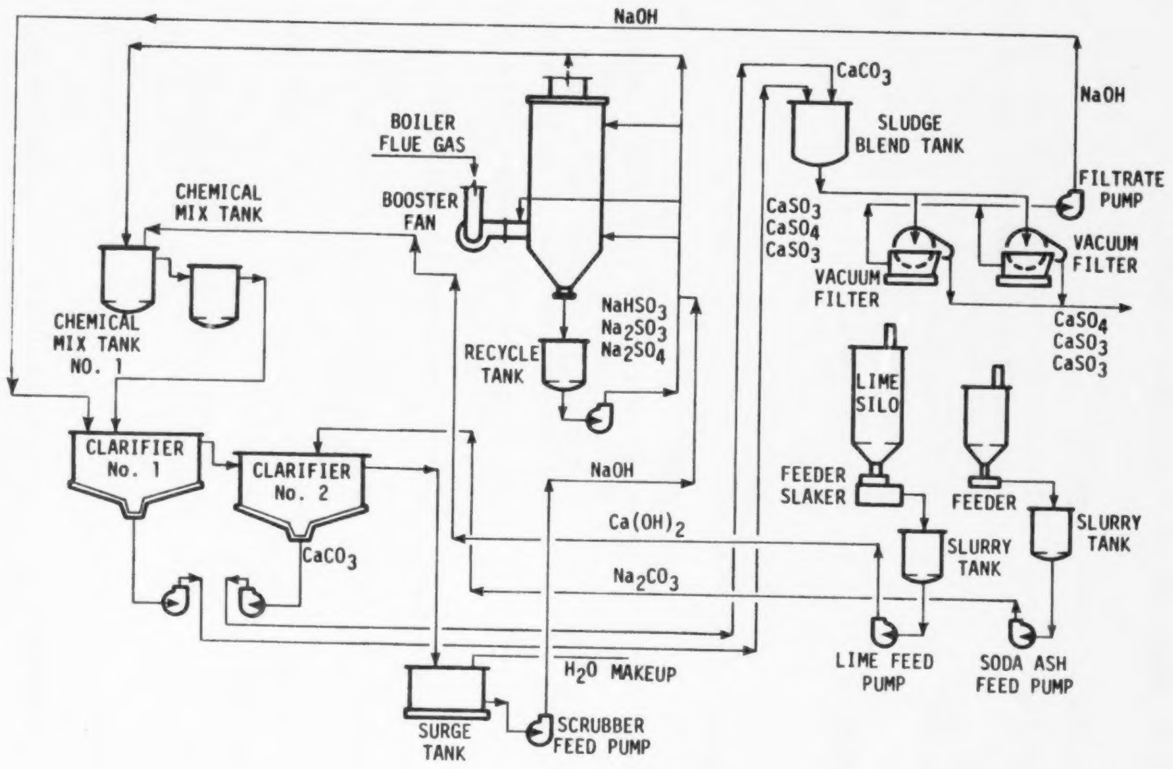

Figure 3-40. Parma double alkali scrubbing system: schematic and simplified process flow diagram.

additional reaction and solids separation. Liquid effluent is then pumped to the second clarifier, where it is softened by addition of $NaCO_3$ only. Solution from this tank is recycled to the scrubber recirculation loop. Underflow from both clarifiers is pumped to the slurry storage tank for batch processing through the vacuum filters. Filtrate is returned to the primary clarifier for recovery of NaOH.

Pertinent design data are given in Table 3-20.

Distinguishing Features - This installation is characterized by:

- Operation in the dilute mode (0.10 normal or less) with lime regeneration and carbonate softening.
- Simultaneous particulate and SO_2 control.
- Firing with coal of 1.5 to 3.0 percent sulfur.

Operating History - The system has performed well with regard to SO_2 removal. Results of a 1-week test in 1974 indicate SO_2 removal efficiencies in the 94- to 99-percent range, with relatively low inlet SO_2 levels (600 ppm to 1200 ppm) and high excess air rates.[23] A test was conducted by A. D. Little and General Motors (GM) from August 19, 1974 to May 14, 1976. It consisted of three, 1-month intensive test periods and 18 months of lower-level tests. Removal of SO_2 reflects the variations in operating modes employed by GM during the period, but removal efficiencies were at 90 percent for the viable operating modes.[24] Operation during April and May 1976 was excellent and A. D. Little recommended continued operation in the mode used during this period.

The operability (hours the FGD system was operated/boiler operating hours in a period expressed as a percentage) of the Parma system from August 1974 through April 1977 is presented in Table 3-21. The system's best period of operation was May through August, 1976, when operability

Table 3-20. GM PARMA DOUBLE ALKALI SCRUBBING SYSTEM DESIGN, OPERATING, AND PERFORMANCE CHARACTERISTICS[24]

Application	Four coal-fired stoker boilers
Fuel, characteristics	Coal: 25,600 to 31,400 J/g (11,000 to 13,500 Btu/lb) 1.5 to 3.0 percent sulfur
System design	Four three-stage multiventuri flexitray scrubber modules
Process/mode	Dilute active alkali
Pressure drop	25 to 33 cm H_2O (10 to 13 in. H_2O)
Status	Operational
Start-up date	3/74
Inlet gas conditions:	
Flow rate	30.9 m^3/s @ 27°C (65,500 acfm @ 80°F)
SO_2	800 to 1300 ppm
O_2	N/A[a]
Particulate	0.7 g/m^3 (0.3 gr/scf)(dry)
SO_2 outlet concentration	20 to 130 ppm
SO_2 removal efficiency	90 to 99 percent
Scrubbing solution characteristics:	
pH	5.5 to 7.5
Total sodium	0.58 to 0.96 $\bar{M}$
Active sodium	0.087 to 0.13 $\bar{M}$
Calcium ion	305 to 490 ppm
Soda ash makeup	>0.1[b] mole/mole SO_2 removal
Lime utilization	1.32 to 1.90 mole/mole S in cake
Filter cake production rate	
Filter cake solids	40 to 55 percent
Filter cake disposal	Offsite landfill

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[a] N/A = not available.

[b] Nonsteady-state operations resulted in makeup rates of 0.028 to 0.05 moles Na_2/mole SO_2 removed.

unit with feed and recycle streams added at the top. The absorption trays (Koch) have movable bubble caps for adjustment to variations in gas flow. Pressure drop through the absorber is designed at 19cm (7.5 in) H_2O and the maximum L/G ratio is 2.7 l/m^3 (20 gal/1000 CF). The liquid stream is composed of about 20 percent fresh feed and 80 percent recycle.

In the initial mode of operation, a side stream from the scrubber recirculation loop was constantly fed to a mix tank, where $CaCO_3$ slurry from the softener clarifier was added for neutralization. Effluent from this tank was pumped to another back-mix reactor, where lime slurry was added for regeneration of caustic. The regenerated solution, with a high concentration of fly ash and calcium precipitates, flowed to two reactor clarifiers in series. In the first clarifier, the solution went through additional reaction and solids separation. Liquid effluent was pumped to the second clarifier where it was softened by the addition of Na_2CO_3 and CO_2. Solution from this tank was recycled to the scrubbers as fresh feed; underflow from the primary clarifier was pumped to a slurry storage tank for batch processing through a vacuum filter; and cake from the filters was hauled to a landfill for disposal. Filtrate was returned to the primary back-mix reactor for recovery of NaOH.

Several changes were made in this operational mode that resulted in a more successful operation. In the modified system, a side stream from the scrubber recirculation loop is constantly fed to the back-mix reactor, where lime slurry is fed for regeneration of caustic. Effluent from this tank is fed to another back-mix reactor for further reaction. The regenerated solution flows to two reactor clarifiers in series. In the first clarifier, the solution goes through

Table 3-21. PARMA DOUBLE ALKALI SCRUBBING SYSTEM PERFORMANCE HISTORY: OPERATION TIME, OPERABILITY, AND COMMENTS$^+$

Month	Operation hours		Operability %	Comments
	Boiler 1 2 3 4	FGD system 1 2 3 4	1 2 3 4	
Aug. 74	555	475	86	Units 1 and 3 were required for service. No. 3 scrubber module was unavailable for approximately 25 hours.
Sep. 74	1190	1015	85	All units were in service during the month. Total available time was approximately 40 hours.
Oct. 74	1355	100	7	Units 1, 2, and 4 were in service during the month. No. 1 scrubber module was the only module in service, operating for approximately 100 hours.
Nov. 74	1245	150	12	All units were required for service. No. 1 scrubber module was the only module in service, operating for approximately 150 hours.
Dec. 74	1495	360	24	Units 1, 2, and 4 were required for service. No. 1 module was the only module in service, operating for approximately 360 hours. Total scrubber system unavailable time was only 70 hours.
Jan. 75	1825	505	28	All units were required for service. Total scrubber system unavailable time was approximately 245 hours.
Feb. 75	1595	455	29	All units and scrubbers were in service. Total scrubber unavailable time amounted to approximately 230 hours.
Mar. 75	1105	280	25	All units were required for service. Scrubber No. 4 was not put in service, although it was available for most of the month.
Apr. 75	1610	440	27	All units were in service. Scrubbers 1 and 4 operated during the period. Total system unavailable time was approximately 40 hours.
May 75	1190	220	18	All units were required for service. Scrubber modules 1, 3, and 4 were operated during the month. Approximately 90 hours of scrubber unavailability time were recorded during the month.
Jun. 75				Recent operation has been intermittent. Tests have been conducted for analysis of particulate loading at the boiler outlet. The system was restarted in May, but was shut down because of a plugged chemical feed line to clarifier. After restart, around June 10, the system was shut down because of similar plugging in another part of the line. Unit ran for about 2 weeks in June.

Table 3-21 (continued). PARMA DOUBLE ALKALI SCRUBBING SYSTEM

PERFORMANCE HISTORY: OPERATION TIME, OPERABILITY, AND COMMENTS

Month	Operation hours: Boiler 1 2 3 4	Operation hours: FGD system 1 2 3 4	Operability % 1 2 3 4	Comments
Jul. 75 Aug. 75				FGD system was down in July and August for replacement of gravity flow lines by an open flume.
Sep. 75 Oct. 75	2331	1848	79	FGD system restarted September 8, and operated during this period at an availability factor of 80 percent. (Note: The figures given for September and October represent the operation hours from September 8 to November 9.)
Nov. 75 Dec. 75	2135	1250	59	The FGD system operated during the report period (November-December), except for a scheduled holiday shutdown from December 23 to January 4, 1976. Because of problems with solids and solids carryover, GM conducted steam tests during December to determine whether solids carryover is due to a high solids recirculation rate. The FGD system was down most of December because of these tests, which caused the low operability factor.
Jan. 76 Feb. 76				GM has found solids concentration in the clarifier too high for efficient system operation. A. D. Little is scheduled to test the system in April as part of an EPA evaluation program.
Mar. 76	1379	240	7	The low operability index during March resulted from extensive modifications performed on the system by GM. During this period, the scrubber blowdown and scrubber flow indication system was revised.
Apr. 76	1084	847	78	From April 19 to the end of the month the system was 100 percent operable. During this period, all system modifications were completed and all major problem areas corrected.
May 76 Jun. 76	1149 924	1042 816	91 88 87	No major problems were encountered during the May-June reporting period. One system upset occurred during the period because of an operator error. Boiler hours were low in June because of low process demand. The scrubbing system characterization study was concluded on May 14, 1976.

Table 3-21 (continued). PARMA DOUBLE ALKALI SCRUBBING SYSTEM

PERFORMANCE HISTORY: OPERATION TIME, OPERABILITY, AND COMMENTS

Month	Operation hours												Comments
	Boiler				FGD system				Operability %				
	1	2	3	4	1	2	3	4	1	2	3	4	
Jul. 76		599				599				100			The boiler demand was very light during the report months (July-August), requiring only the boiler's being continuously on line. The system was down for one week because of the annual plant inventory. Modifications to the operation of the gas scrubbing system included: ° Elimination of polymeric addition. ° Elimination of sludge blanket for filtration. ° High-percentage sulfate concentration in the filter cake. ° Incorporation of a pH controller in the chemical mix tank regulating the lime feed into the system. pH control should reduce the current stoichiometric requirements from the 1.25 to 1.50 range down to 1.10 to 1.20.
Aug. 76		715				715				100			
Sep. 76		809				223				40			Two boilers and their corresponding scrubber modules were in service during the month. Operation time was logged primarily in the initial part of the month. A number of problems were encountered, forcing a shutdown throughout the remainder of the month. These included: Solids deposition and plugging of the chemical tanks and reactors, requiring shutdown and cleanout. Agitator problems. Boiler problems. Replacement of demister pads.
Oct. 76		1174				734				63			The system was returned to service the second week in October. Two boilers and their corresponding scrubbers were put back in the gas stream. Problems encountered stemmed primarily from a mechanical collector malfunction, resulting in excessive dust loadings in the modules and eventually causing widespread plugging of the lower bubble-cape trays. Shutdown and cleanout of the modules plus repairs to the mechanical collector were required. System restarted November 8. Major modifications on the double alkali system during the period included the installation of a Great Lakes pH monitor and control unit for the regulation of lime feed to the regeneration reactor-clarifier unit. This unit employs a digital readout and is scheduled for operation in November. Running a seeding line back to the reactor-clarifier unit resulted in superior crystal growth by the calcium sulfite-sulfate salts.

Table 3-21 (continued). PARMA DOUBLE ALKALI SCRUBBING SYSTEM

PERFORMANCE HISTORY: OPERATION TIME, OPERABILITY, AND COMMENTS

Month	Operation hours								Operability %				Comments
	Boiler				FGD system								
	1	2	3	4	1	2	3	4	1	2	3	4	
Nov. 76			1321				787				60		The FGD figures reported for the month of December are somewhat distorted, because only UAW personnel can operate the system; therefore, the scrubbers could not be operated during the Christmas holiday. During the latter part of November and early December, the scrubbers were taken out of the flue gas path, because of continued solids buildup in the mix tank. In addition, some piping arrangements were modified and the gear reducer in the mix tank had to be replaced.
Dec. 76			1559				349				22		
Jan. 77			1852				608				33		Lost an agitator on the No. 2 chemical mix tank. Low operability factors for the winter months due primarily to severe weather conditions and plant curtailments.
Feb. 77			1462				892				61		
Mar. 77			1476				1278				87		March and April operations were very satisfactory. SO_2 removal has been measured well above the 90-percent level. Particulate removal has been below par because of solids carryover.
Apr. 77			1300				1158				89		

+ Based on reference 23 and personnel communication between PEDCo and G.M. personnel.

averaged 94 percent. The GM Parma plant has several unique characteristics that affect operability. Each boiler is equipped with its own separate scrubbing module with no provision for crossflow between modules. The plant is not needed during the summer months, because operations are shutdown during automobile model changeover changes and because there is no need for heating. United Automobile Workers personnel operate the scrubber plant, which precludes operation of the scrubbers when they are not on the site. The GM plant is a developmental system, and as such is subject to modifications. Many of the low operability periods were due to mechanical outages, or outages for modifications to accomodate and test new modes of operation.

Operational Problems/Solutions[25] - Major problems encountered at the Parma plant were:

1) Scaling in the scrubber loop
2) Periodic excess entrainment
3) Excessive lime consumption
4) Excessive soda ash consumption
5) Low solids content in sludge

Solutions to these problems are presented below:

1) Scaling in the scrubber loop - During the test program at the Parma plant, scaling occurred in the scrubber loop under two sets of circumstances. In the first instance, calcium carbonate scale formed on the top tray of the scrubber, where fresh scrubber feed mixed with a recycle stream. The high pH levels on this tray and the lack of mixing resulted in the absorption of CO_2 from the flue gas and formation of $CaCO_3$. General Motors attempted to eliminate the problem by introducing fresh scrubber feed to the recycle tank for better pH control. In this scheme, calcium

sulfite scale formed on the scubber internals. This was caused by the high pH in the recycle tank, high calcium ion levels, and low sodium ion strength. The company reduced this scaling by installing a line to mix regenerated liquor with scrubber recycle liquor at the entrance to the top tray. The major difference between this and the first mode of operation is that in the new mode the recycle stream and the fresh regenerated slurry feed are mixed entirely outside the scrubber in a pipe. This assures good mixing and eliminates localized high pH.

2) Periodic excess entrainment - Although entrainment of scrubber effluent is within design specifications, at times it can contribute significantly to particulate loading. This was the result of poor gas distribution to the demister, rendering a portion of it ineffective. Although no remedial measures were attempted by GM, it was discovered that replacement of the original stainless steel wire mesh entrainment separators with polypropylene units virtually eliminated visual evidence of entrainment problems.

3) Excessive lime consumption - Lime consumption has been significantly higher than what was expected in theory. It has been improved, however, from an initial stoichiometry value of 1.92 moles Ca per mole SO_2 removed to a value of 1.32 moles Ca per mole SO_2 removed. Problems occurred because of difficulties with the automatic lime feed system which forced adoption of manual control, and resulted in the low sodium level in the system. The scrubber liquor exhibited an Na2SO4 concentration of 0.33 M, at which the solubility of $[OH^-]$ is less than 0.1 $\underline{M}$. The lime feed rate was controlled on the basis of a target hydroxide level of 0.1 $\underline{M}[OH^-]$, resulting in overfeeding of lime. Lime consumption was improved by adjusting the feed rate to give 0.08 to 0.09 $\underline{M}$

$[OH^-]$ in chemical mix tank No. 2, compared with the previous value of 0.1 M, and by installing a flow orifice to measure the scrubber feed rate accurately.

4) Excessive soda ash consumption - In the early stages of operation, solubles loss was reported in excess of 0.1 moles Na_2 per mole SO_2 removed. This was caused by failure to wash the filter cake, by high liquor content in the filter cake, and by high sodium in the liquor. Washing the filter cake has eliminated this problem to the point that soda ash makeup is now determined by the minimum amount required for the softening step. This amount has not yet been determined, but with softening requirements corresponding to Ca^{++} reduction of 150 ppm (to 650 ppm), the stoichiometry is 0.05 moles Na_2 per mole SO_2 removed.

5) Low solids content in sludge - Excessive lime consumption and low oxidation rates resulted in a solids content in the filter cake of 39 to 46 percent. This was improved to 56 percent in the latest test runs using cake washing and the maintenance of low excess lime feed.

Summary - Operation of the Parma facility has shown the viability of the double alkali process for removal of sulfur dioxide. Removals of 90 percent or better have been obtained on coals ranging from 1.5 to 3.3 percent sulfur [25,600 to 31,400 J/g (11,000 to 13,500 Btu/lb)].[26] Several different operating modes have been investigated and significant improvements have been obtained in both process and mechanical performance. Although it has yet to be proved over an extended test period, it is believed that in the latest operational mode the system is capable of long-term reliability.[27]

3.3.2.2 Caterpillar Tractor - Joliet Plant

System Description[28,29,30] - Figure 3-41 is a simplified process flow diagram for the Joliet double alkali FGD system. Flue gases from two coal-fired stoker boilers are routed to two Zurn Dustraxtor scrubbers. Here the gases are continuously contacted with a dilute aqueous sodium hydroxide/sulfite/bisulfite solution. The pressure drop through the scrubber varies from 25 to 30 cm H_2O (10 to 12 in. H_2O), with a fresh feed L/G ratio of 8 l/m^3 (60 gal/1000 acf). A side stream of solution is withdrawn from the scrubber to an external mix tank. Here, slaked quick lime is added for neutralization and regeneration of active alkali. The solution is drawn to a second mix tank, where the reaction continues. Insoluble calcium salts precipitate as the $Ca(OH)_2$ reacts with the sodium salts. Underflow from the second mix tank is pumped to a vacuum filter and the calcium salts are dewatered to about 65 percent solids. This slurry is then disposed of in an offsite landfill. The vacuum filter is washed with fresh water to recover as much of the soluble sodium as possible. Filtrate from the vacuum filter, which contains a high concentration of calcium ions, is recycled back to a clarifier. The clarifier also receives the regenerated solution overflow from the thickener. Sodium carbonate is added to the clarifier to "soften" the solution by reacting with the calcium ion to form calcium carbonate precipitate. Underflow from the clarifier is pumped to the vacuum filter for dewatering of the calcium carbonate before disposal. Overflow from the clarifier, regenerated active alkali solution, is recycled to the scrubbers. Pertinent design and operating data for the Joliet plant are presented in Table 3-22.

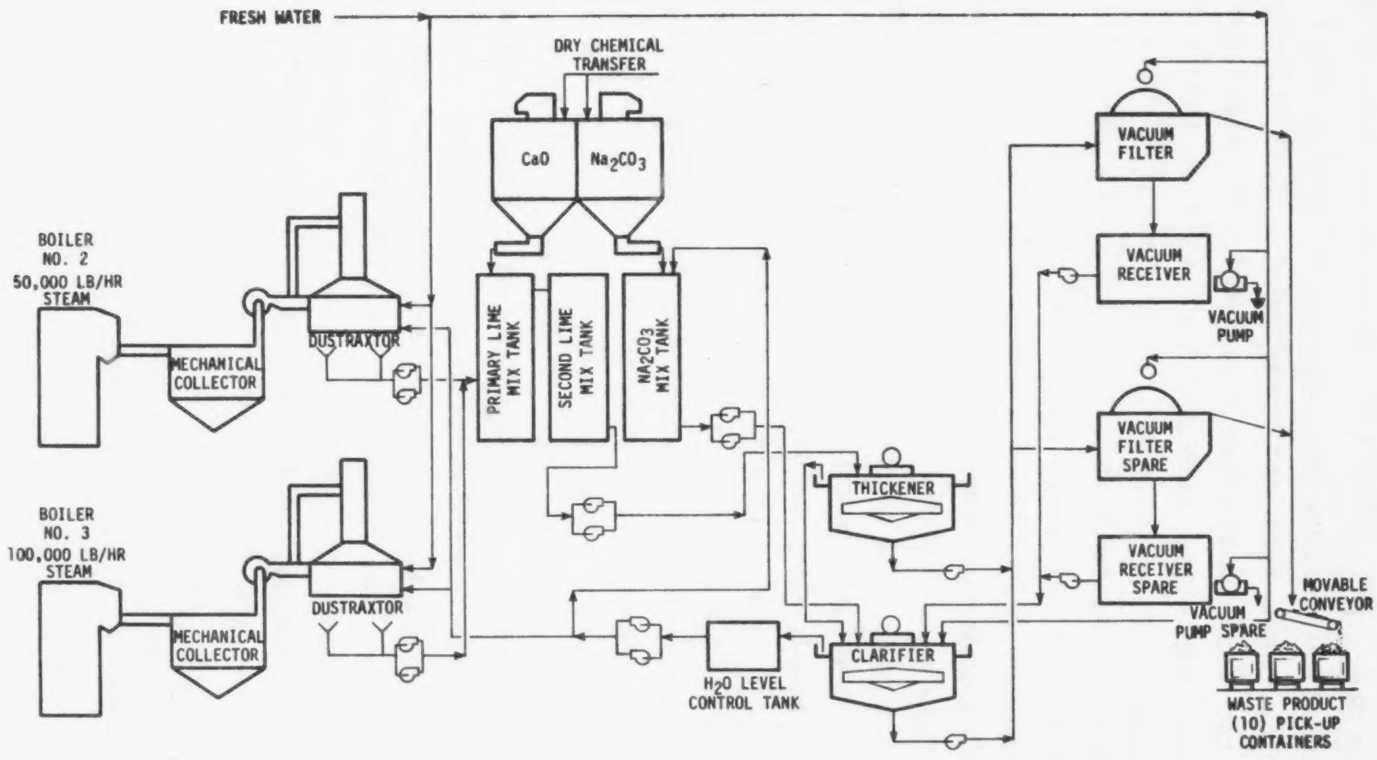

Figure 3-41. Joliet double alkali scrubbing system: schematic and simplified process flow diagram.

Table 3-22. JOLIET DOUBLE ALKALI SCRUBBING SYSTEM DESIGN, OPERATING, AND PERFORMANCE CHARACTERISTICS[31]

Application	Two coal-fired stoker boilers
Fuel, design characteristics	Coal: 27,600 J/g (11,869 Btu/lb) 4 percent sulfur
System design	Two dustraxtor scrubber modules
Status	Operational
Start-up date	9/74
Design inlet gas conditions:	
Flow rate	48.8 m^3/s @ 182°C (103,500 acfm @ 360°F)
SO_2	2325 ppm
SO_2 outlet concentration	115 to 350 ppm
Actual SO_2 removal efficiency	85 to 95 percent
Soda ash consumption	35.9 g/s (285 lb/hr)
Lime consumption	131 g/s (1,040 lb/hr)
Water makeup	0.038 l/s (36 gph)
Power consumption	6.85 kWh/hr
WAste disposal	655 g/s (5,200 lb/hr)
Filter cake, percentage of solids	50 to 60 percent
Total operation time	8600 hr
System availability	90 to 100 percent

Distinguishing Features - This installation is characterized by:

- Operation in the dilute mode with lime regeneration and carbonate softening.
- Simultaneous particulate and SO_2 removal in multitube impingement/entrainment scrubber.
- Firing coal of about 4.0-percent sulfur content.

Operating History - The system has achieved excellent SO_2 removal efficiencies of between 85 and 95 percent under various operating conditions.[32] Sulfur dioxide inlet concentrations are high, about 2300 ppm. The system was designed to attain an emission level of 0.86 g SO_2/MJ (1.9 lb $SO_2/10^6$ Btu) (75% SO_2 removal), but has consistently performed much better than design.

The operability of the FGD system has been improving steadily. Process availability for the period October 24, 1975, through June 1976 has been 100 percent.[33] Most problems at the Joliet plant are mechanical; the majority are solved while still on-stream or during scheduled shutdowns resulting. As a consequence, there have been few forced outages.

Operational Problems/Solutions[34,35] - Problems encountered at the Joliet plant included:

1) Unreliability of the chemical feed system
2) Inoperability of the SO_2 monitors
3) Excessive water consumption
4) Feed water pump impeller cavitation
5) Poor vacuum filter belt life

Reasons for these problems and their solutions are presented below:

1) Unreliability of the chemical feed system - Two problems arose with the lime and soda ash supply systems. The first was the failure of electronic components in the feed control system, resulting in inaccurate feed rates. This was alleviated by replacement of the faulty electronic components. The second problem was the formation of lumps inside the storage silos. They were caused by moisture condensation on the inside walls and leakage of rain water into the silos. The lumps also caused inaccurate feed rates. The problem was solved by the installation of a dry air purge system on each silo.

2) Inoperability of SO_2 monitors - The system employs several monitors, in the inlet duct and in the stack, to measure SO_2 concentrations. They experienced problems caused by pluggage of the sample probes. The difficulties were removed by increasing the pressure of the blow-back air used to clean the probes, and by the installation of shields around the probes.

3 Excessive water consumption - During early operation, the scrubber did not achieve closed-loop operation with respect to water, because of the addition of excess water through the process pump seals; from vacuum filter wash sprays; and as a result of the low water evaporation rates in the scrubber when it was under low boiler loads. Closed-loop operation was obtained by installation of rotameters on the water supply lines of the process pump seals, and by using clarified scrubber feed solution for the vacuum filter belt wash. A seal water recirculation system was also installed on the vacuum pumps.

4) Feed water pump impeller cavitation - Insufficient suction head in the surge tank caused cavitation, which in turn led to impeller damage. This was solved by increasing

the surge tank height to raise the suction head at the feed water pumps.

5) Poor vacuum filter belt life - Originally the life of the multifilament polypropylene filter belt was poor, because of the accumulation of solids in the interstices of the cloth. The pressure of the belt wash sprays was increased; and this resulted in extended belt life. Other materials and weaves are being tested to see if belt life can be increased even further.

Summary - The Joliet facility has demonstrated the viability of a double alkali FGD system for control of SO_2 on coal-fired boilers. Sulfur dioxide removals of 85 to 95 percent have been demonstrated on high-sulfur coal together with very good operability. The problems encountered, which were mechanical and generally associated with ancillary hardware, have been effectively solved.

3.3.2.3 Caterpillar Tractor - Mossville Plant[36,37] - The double alkali FGD system at the Mossville Engine Plant of Caterpillar Tractor Co. is the largest installed to date by FMC Corporation. It went into commercial service in October 1975. The system employs a dual-throat, venturi-type scrubber to remove SO_2 and particulates. The scrubber handles flue gas from four coal-fired stoker boilers at the rate of 113 m^3/s (240,000 acfm). The boilers are rated at 58 kg/s (460,000 lbs/hr) of steam (57 MW equivalent).

The system operates in the concentrated mode (>$0.15\bar{M}$ active alkali) using lime for regeneration of the sodium sulfite scrubbing liquor. Precipitated calcium solids are dewatered in a vacuum filter to about 60 percent solids before disposal in a landfill.

FMC reports very high system availabilities (90%) and 90-percent SO_2 removal efficiencies. The problems experienced at Mossville are related to wear of control valves, filter cloth, and conveyor belts. FMC feels that satisfactory solutions have been reached for all these problems.

3.3.2.4 Firestone - Pottstown Plant

System Description[38,39] - Figure 3-42 is a simplified process flow diagram for the Pottstown plant. A slipstream of flue gas equivalent to about 3 MW is routed to the scrubbing system from a pulverized-coal boiler (formerly burning 2-percent sulfur oil, but being converted to burn 3.0-percent-sulfur coal). The flue gas is contacted with the scrubbing liquor in a dual-throat venturi scrubber. It then moves to a cyclone separator, which collects the scrubbing liquor and discharges it to a recirculation tank. Cleaned flue gas then passes out the top of the separator, through a mist eliminator, and on to the stack. A side stream is drawn off the recycle tank and fed to a reaction tank. Here, slaked lime is added to regenerate the sodium sulfite, bringing about precipitation of calcium solids. The solution is then pumped to a thickener, where the calcium solids are settled and the thickened slurry passed to a rotary drum filter. The slurry is dewatered to a cake of about 55 percent solids and disposed of in a landfill. Supernatant from the thickener and the filter is pumped back to the recycle tank. At this stage, sodium carbonate is added to make up for sodium lost in the filter cake. Pertinent design and operating data for the Pottstown Plant are given in Table 3-23.

Distinguishing Features - This installation is characterized by:

- Operation in the concentrated mode (3 M active alkali) with lime regeneration and carbonate softening.

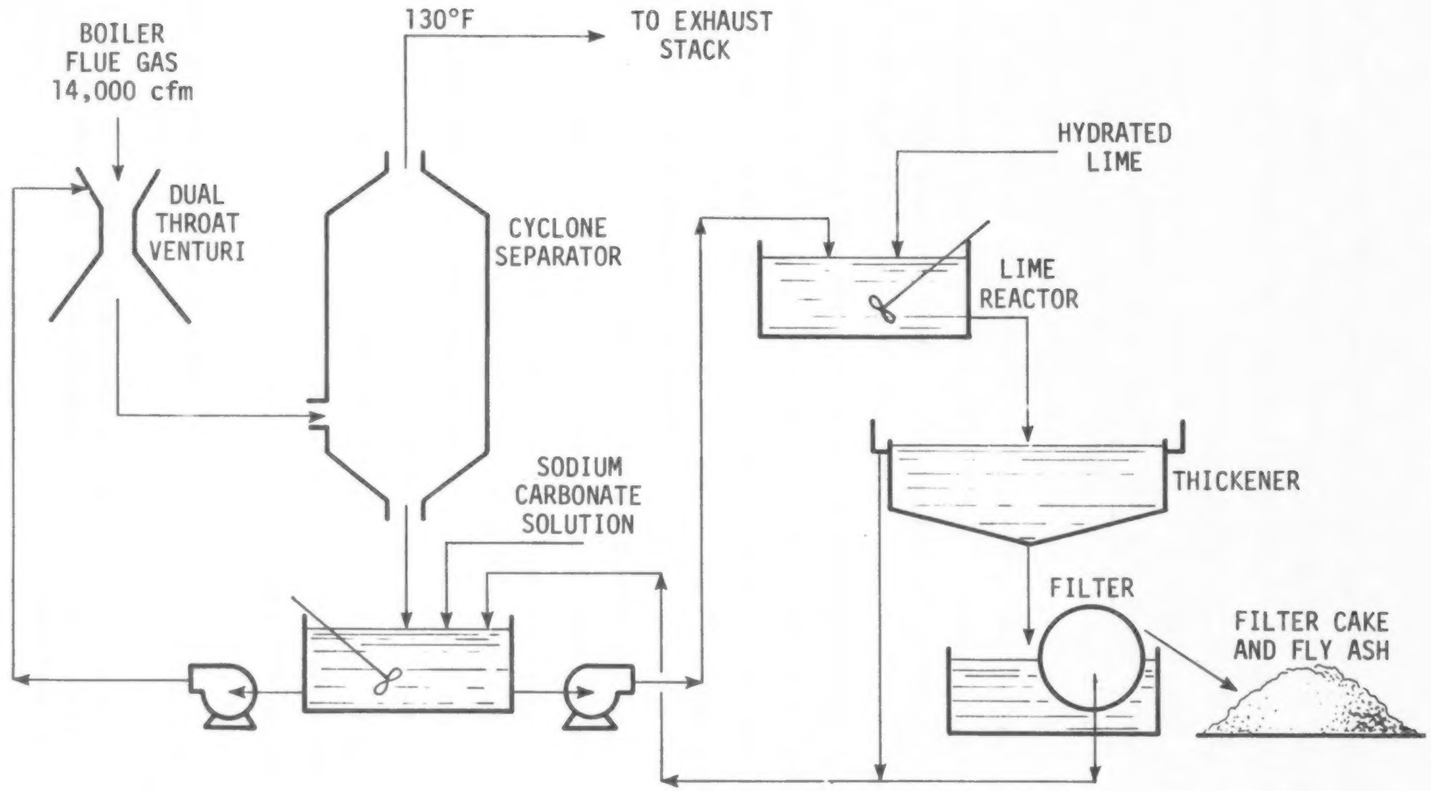

Figure 3-42. Pottstown double alkali demonstration plant: schematic and simplified process flow diagram.[40]

Table 3-23. FIRESTONE POTTSTOWN DOUBLE ALKALI SCRUBBING SYSTEM: DESIGN, OPERATING, AND PERFORMANCE CHARACTERISTICS

Application	Power boiler slipstream (3 MW equivalent)
Fuel, characteristics	Oil, 2 percent sulfur; or coal, 3 percent sulfur
System design	Dual-throat venturi scrubber/ cyclonic separator
Pressure drop	25 to 76 cm H_2O (10 to 30 in. H_2O)
Status	Operational
Start-up date	1/75
Inlet gas conditions:	
Flow rate	6.6 m^3/s (14,000 acfm)
SO_2	1,000 ppm
SO_2 outlet concentration	100 ppm
Actual SO_2 removal efficiency[44]	90 percent
Scrubbing solution characteristics:	
pH	6.5
Total sodium	3.0 $\bar{M}$
Active sodium	0.2 to 0.3 $\bar{M}$
Soda ash makeup of solution	0.019 m^3/h (5 gph)
Lime utilization	100 percent
Filter cake production rate	23 g/s (180 lb/hr)
Filter cake solids	55 percent
Filter cake disposal	Landfill
Actual system availability[45]	94 percent

- Simultaneous particulate and SO_2 removal.
- Operation on 2-percent-sulfur oil or 3-percent-sulfur coal.

Operating History - The system has exhibited excellent SO_2 removal efficiencies of 90 percent on high-sulfur oil, but no data are available for its performance on coal.[41] It has also achieved a very high availability: 99 percent for the first 12 months of operation.[42,43] Most downtime periods were due to mechanical component failure or to maintenance, and not to unwanted chemical changes or side reactions. No scaling problems have been experienced. Table 3-24 presents a summary of the causes for system downtime during 1976.

Table 3-24. AVAILABILITY OF THE FIRESTONE POTTSTOWN DOUBLE ALKALI SCRUBBING SYSTEM: 1976 OPERATING HISTORY

Item responsible for outage	Percentage of total downtime	Percentage of operating period (year)
Thickener plugging	20.7	1.35
Pumps	16.1	1.04
Cake conveyor	16.0	1.04
Fan	15.4	1.00
Lime feeder	13.4	0.87
Spray nozzles	12.4	0.80
Instrumentation	4.1	0.27
Control valves	1.9	0.13
Total	100.0	6.50

Total availability for period = 93.5 percent

An ongoing test program is being conducted at the Pottstown facility. Its goals include the determination of corrosion characteristics of the various components; optimization of filter and thickener systems; and performance of coal-firing versus oil-firing.

Operation Problems/Solutions[46] - Major problem areas encountered at the Pottstown Plant included:

1) Thickener plugging
2) Failure of recirculating pump
3) Fan outage

Reasons for these problems and their solutions are given below:

1) Thickener plugging - the single largest outage was caused by plugging of the thickener underflow pipe after a plastic beaker had been accidently dropped into the unit. No mechanical or chemical upsets occurred in the unit during the first year.

2) Failure of recirculation pump - recycle pump failure, due to separation of the pump lining from the pump housing, was caused by improper maintenance of the pump seal. A proper maintenance program has eliminated this problem.

3) Fan outage - the fan failed because bearings overheated, which in turn was caused by misalignment of the fan shaft. Proper alignment has alleviated this problem.

Summary - Operation of the Pottstown facility has demonstrated the high efficiency and reliability of a system operating in the concentrated mode. This pilot plant is a scale-up of a smaller pilot plant facility; it has confirmed virtually all the process data and relationships developed in that program.[47] Although the Pottstown facility has demonstrated SO_2 removal efficiencies of 90 percent and an availability of about 94 percent on oil-firing, future testing will provide an indication of performance on coal-firing.

3.3.2.5 Utah Power & Light Co. - Gadsby Station

System Description[48,49] - Figure 3-43 is a simplified process flow diagram for the double alkali pilot plant installed by Envirotech on Unit 3 at the Gadsby Station.

A simplified process flow diagram of a full scale double alkali system now under construction at the 575-MW Newton Station of Central Illinois Public Service Co., is provided in Figure 3-44. At the Gadsby Station, a slipstream of 1.2 actual m^3/s (2500 acfm) from the 100-MW coal-fired unit was routed to the pilot plant. The flue gases are quenched in a venturi quench section and fly ash, chlorides, and particulate matter are removed. The flue gas then flows countercurrent to an aqueous solution of sodium hydroxide/-sulfite/bisulfite in a mobile-bed scrubber with two trays of polysphere balls. Entrained liquid in the flue gas is removed in a mist eliminator before discharge of clean flue gas to the stack. The spent scrubbing solution is collected in a recycle tank, where atmospheric air is bubbled through the solution to oxidize the sodium sulfite to sodium sulfate. A side stream from the recycle tank is pumped to two reaction tanks and slaked lime is added for regeneration of sodium hydroxide. The solution then passes to a reactor-clarifier to settle the precipitated calcium sulfate solids. Underflow from the reactor-clarifier is sent to a thickener, in which calcium solids are dewatered.

The thickened slurry (40% solids) is then pumped to a rotary vacuum filter so the slurry can be further dewatered to a calcium sulfate cake (70-80% solids). The cake is washed to recover any soluble sodium still present in it, and the wash solution is pumped back to the reactor-clarifier.

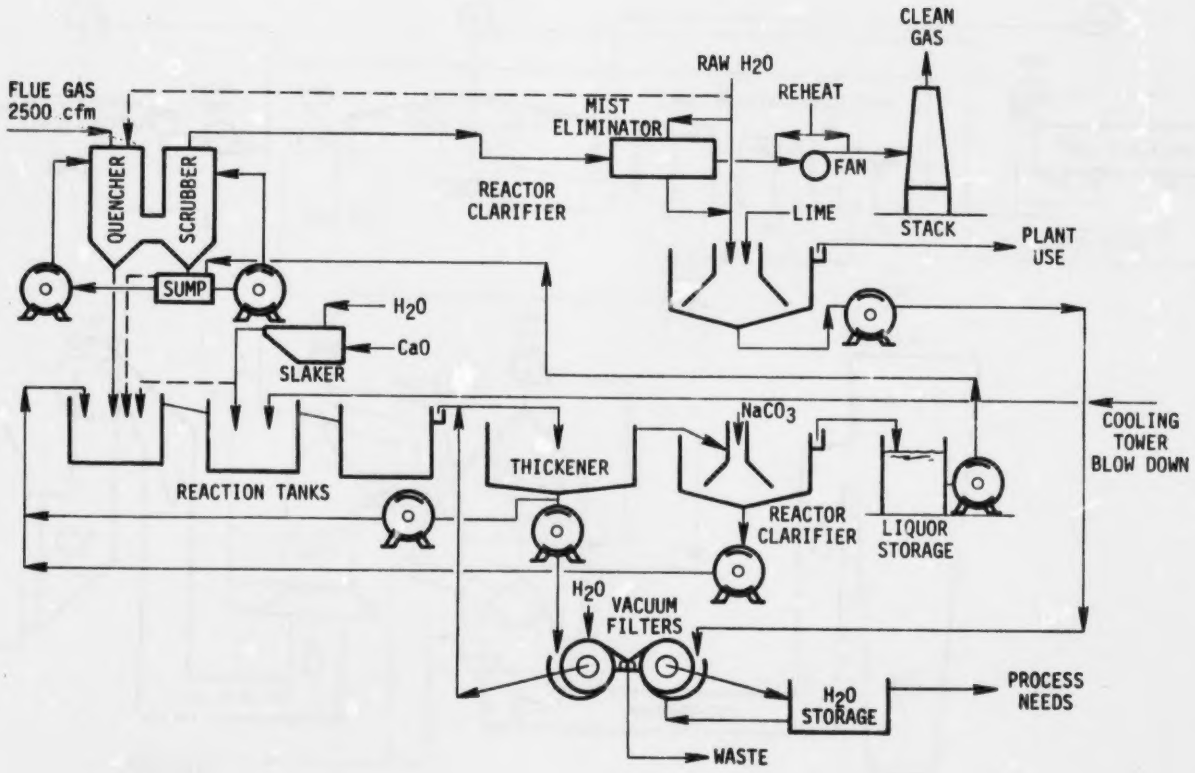

Figure 3-43. Envirotech Gadsby double alkali pilot plant: schematic and simplified process flow diagram.

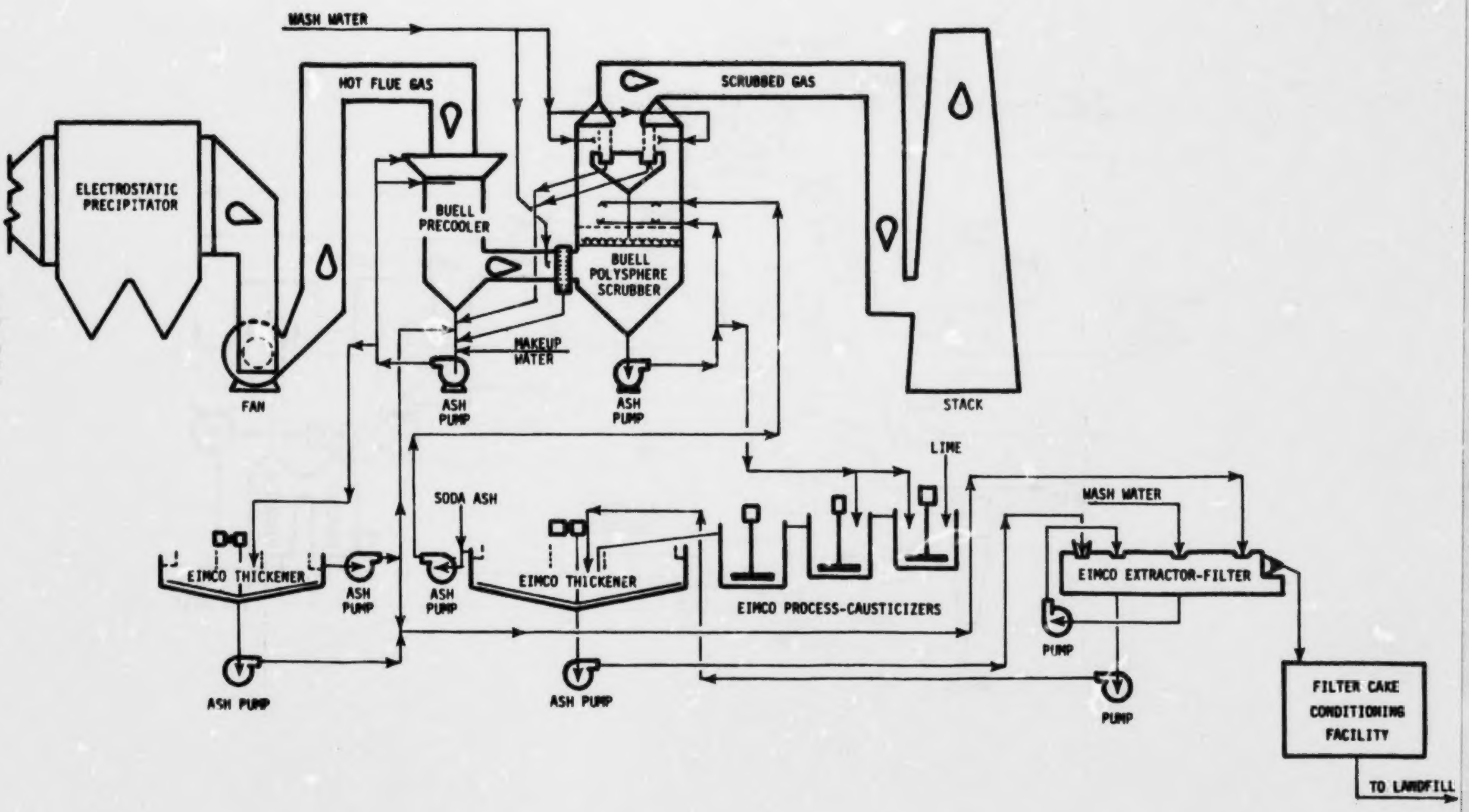

Figure 3-44. Double alkali system at Newton Station.

Supernatant from the reactor-clarifier and the thickener is pumped to a second reactor-clarifier. Soda ash (sodium carbonate) is added to this reactor-clarifier as a softener. Excess calcium ions are precipitated as calcium carbonate, which is then pumped to the first reaction tank. Overflow from the reactor-clarifier is pumped to the recycle tank.

This installation is a pilot plant, and as such has been tested in a number of different configurations. In addition to the polysphere scrubber, a venturi scrubber has been tested. Two types of mist eliminators have been used, a spin vane in series with a cyclone, and the Euroform Chevron type. Most of the testing was done in a dilute mode, but some was also performed in the concentrated mode. Pertinent design and operating data are given in Table 3-25.

Distinguishing Features - This installation is characterized by:

- ° Operation in both the dilute mode and the concentrated mode with lime regeneration and carbonate softening.
- ° Simultaneous particulate and SO_2 removal.
- ° Firing with 0.4-percent-sulfur coal.

Operating History - The scrubbing system has performed well with respect to SO_2 removal. Various modes of operation were tested using two types of absorbers. With the polysphere absorber, SO_2 removals of 90 percent were achieved, giving outlet concentrations of 15 to 40 ppm SO_2.[51] With the venturi absorber, efficiencies ranged from 80 to 85 percent SO_2 removal.

With the exception of the first three-month operating period, during which some gypsum scaling problems were encountered, dilute mode operations were conducted for

Table 3-25. ENVIROTECH/GADSBY PILOT PLANT TEST PROGRAM: DILUTE MODE PERFORMANCE AND RESULTS[50]

Application	Pulverized coal power boiler
Fuel, characteristics	Coal: 0.4 percent sulfur
Gas inlet conditions:	
Flow rate	1.2 m^3/s @ 104°C (2500 acfm @ 220°F)
SO_2	250 to 1500[a] ppm
O_2	5.0 percent
Particulate	0.5 to 0.7 g/m^3 (0.2 to 0.3 gr/scf)
SO_2 outlet	15 to 40 ppm
Particulate outlet	0.02 to 0.05 g/m^3 (0.01 to 0.02 gr/scf)
SO_2 removal efficiency	90 percent
Particulate removal efficiency	93 to 95 percent
L/G ratio	2.7 l/m^3 (20 gal/1000 acf)
Pressure drop	10 to 25 cm H_2O (4 to 10 in. H_2O)
Recycle liquor composition:	
pH	5.8 to 7.7
Active sodium	0.10 $\bar{M}$
Calcium ion	500 ppm
Sulfate ion	40,000 to 45,000 ppm
Chloride ion	1000 ppm
Lime consumption, percentage stoichiometric	110 to 115 percent
Reactor residence time	120 min
Filter cake composition:	
$CaSO_4 \cdot 2H_2O$	42 to 57 percent
$CaCO_3$	10 to 15 percent
H_2O	30 to 40 percent
Solubles	2 percent
Ash	1 to 2 percent

[a] During later test runs SO_2 was added to the flue gas to simulate higher sulfur conditions.

almost two years without any major problems.[52] No operating problems causing shutdown were experienced between October 1972 and August 1974. It should be noted that for convenience the system was shut down on weekends, but no drainage of solution or cleaning of equipment took place during these shutdowns.

The majority of the testing with the polysphere scrubber was conducted at a total pressure drop of 10 cm H_2O (4 in. H_2O) using two trays of balls. Figure 3-45 shows the performance of a single polysphere tray; SO_2 removal is plotted against L/G ratio in $1/m^3$ (gal/1000 acf) at three different pressure drops, 10, 18, and 25 cm (4, 7, and 10 in.) H_2O and at a pH from 5.8 to 7.1. Figure 3-46 shows a plot of SO_2 removal versus scrubber effluent pH at an L/G ratio of 2.5 $1/m^3$ (19 gal/1000 acf) at a total pressure drop of 10 cm (4 in.) H_2O across two polysphere trays. It can be seen that, generally, two trays operating at a pressure drop of 10 cm (4 in.) H_2O are equivalent to one tray operating at a pressure drop of 18 cm (7 in.) H_2O with L/G ratios of 2.5 $1/m^3$ (19 gal/1000 acf) in each case.

The scrubber/regeneration liquors had a chloride content of about 1000 ppm at steady state. Soluble chloride usually concentrates in wet scrubbing systems, owing to the conversion of coal chloride to HCl, which appears in the flue gas and is absorbed by scrubber liquors. No obvious detrimental effects were noted for chloride levels in the range experienced in this double alkali application.[55]

Operational Problems/Solutions - The major problem at the Gadsby facility was gypsum scaling in the scrubber, caused by excess calcium concentrations in the recycle slurry. In the original operating flow scheme only one reactor-clarifier was used. The scaling problem was solved, therefore, by the addition of a second reactor-clarifier, in

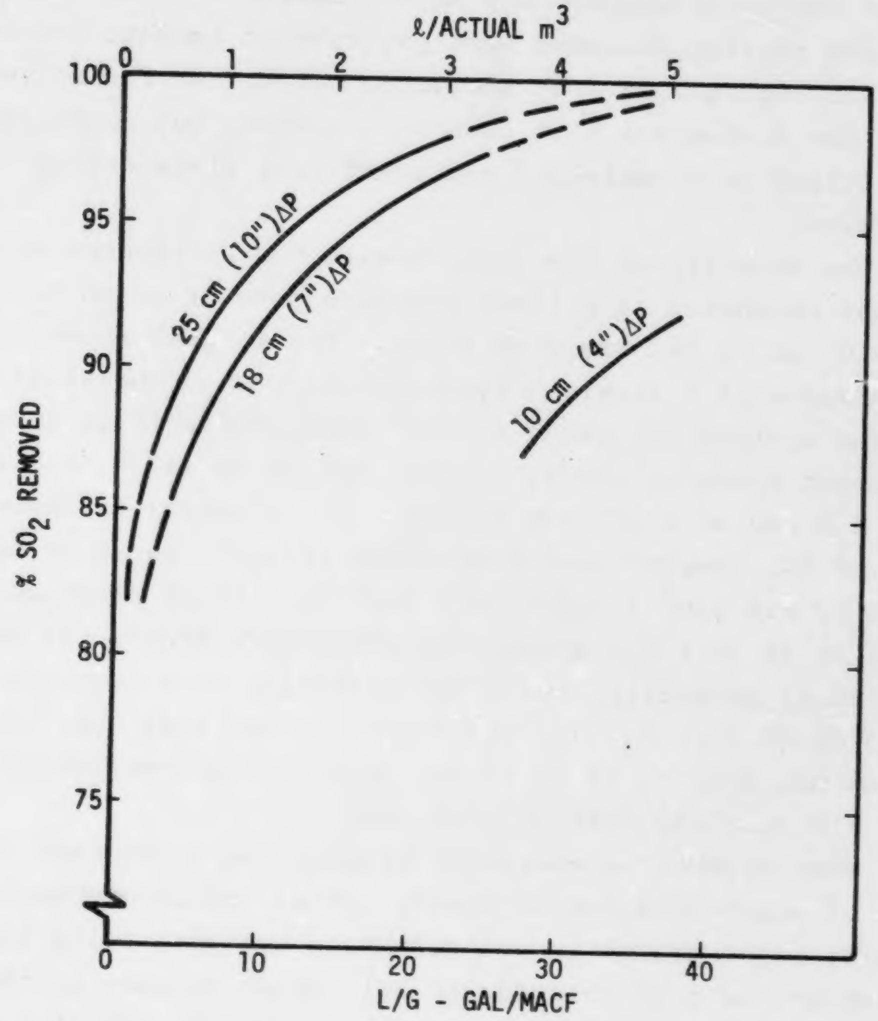

Figure 3-45. SO_2 removal versus L/G ratio for the Envirotech/Gadsby pilot plant with a single stage polysphere absorber.[53]

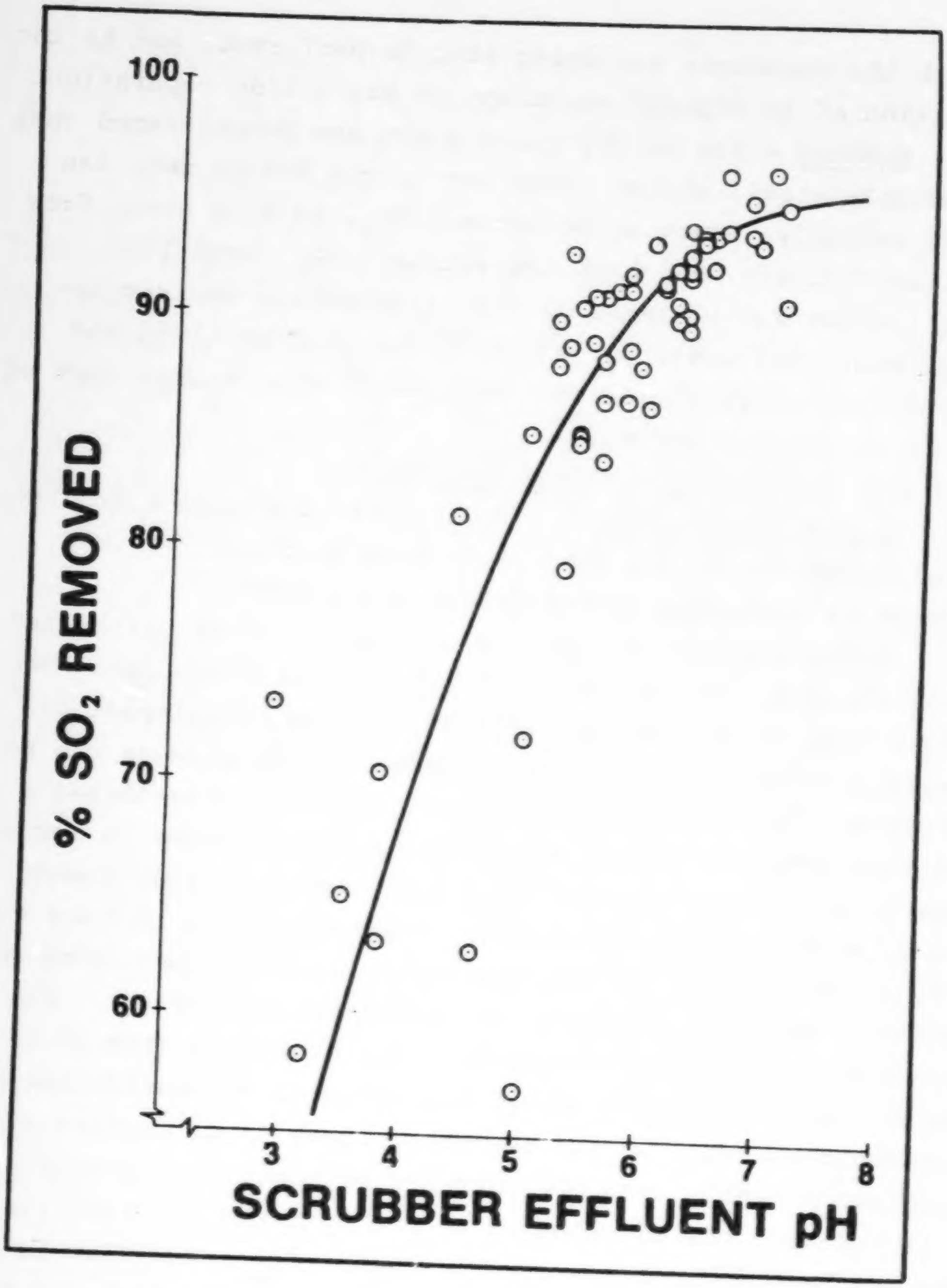

Figure 3-46. SO_2 removal versus scrubber effluent pH for the Envirotech/Gadsby pilot plant at an L/G of 2.5 l/m^3 (19 gal/1000acf) and using a two-stage absorber.[54]

which the carbonate softening step is performed, and by the addition of an organic coagulant to aid solids separation.

Summary - The Gadsby pilot plant has demonstrated that a soluble alkali system operating in the dilute mode can work reliably, removing 90 percent SO_2, on flue gases from utility boilers that burn low-sulfur coal. Good lime utilization was achieved (110 to 115% stoichiometric requirement) and waste solids exhibited good settling and dewatering characteristics, thus enabling an end-product of 60 to 70 percent solids.[56]

3.3.2.6 Gulf Power - Scholz Station

System Description[57] - Figure 3-47 presents a simplified schematic of the CEA/A.D. Little double alkali FGD system as installed on the Scholz Steam Plant.

Fifty percent of the flue gas from a 40-MW pulverized coal-fired boiler is routed to a high-efficiency ESP. The efficiency of the ESP is variable, so that the impact of various levels of particulate matter on the process can be studied. The flue gases then pass to a variable-throat venturi scrubber and an absorption tower arranged in series. The tower can be operated as a spray tower, a tray tower, or as a de-entrainment separator. The venturi can be used for SO_2 absorption or for particulate collection, depending on whether water or absorbent is circulated through it. The pressure drop across the venturi can be varied from 13 to 25 cm (5 to 10 in.) H_2O. After the venturi, saturated flue gas enters the bottom of the absorber, where it is contacted countercurrently with the circulating absorption liquor. Cleaned flue gas is reheated with an externally fired oil burner before discharge to the stack. Spent solution is collected in the internal recycle tank at the bottom of the

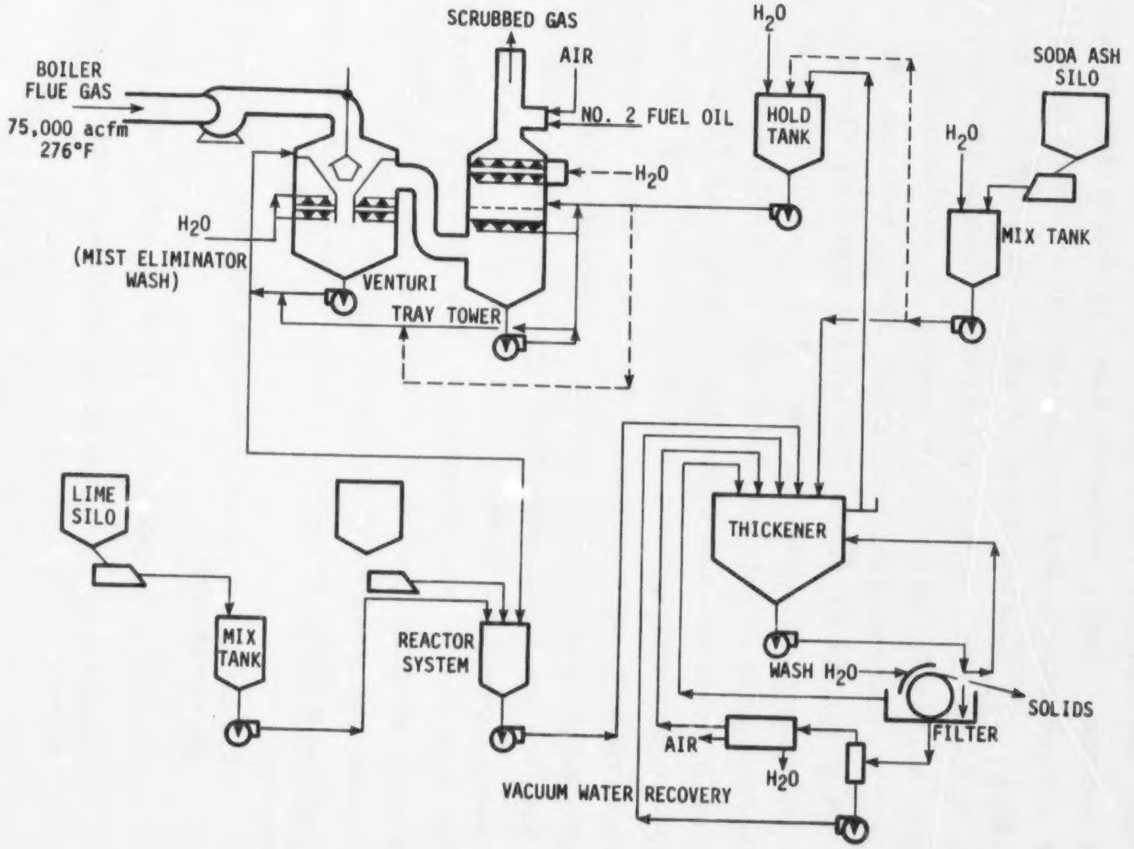

Figure 3-47. Scholz double alkali prototype scrubber: schematic and process flow diagram.

absorber. A bleed from the collected liquor (pH 5 to 7) is sent to the venturi recirculation loop for additional SO_2 absorption. A continuous bleed stream (pH 4.8 to 6.0) is drawn from the venturi recirculation loop and pumped to a reactor system. Here, hydrated lime is added for regeneration of active alkali and precipitation of calcium sulfite and sulfate. Sufficient lime is added to produce a pH of 11.0 to 12.5 in the reactor underflow. The underflow, about 5-percent solids by weight, is pumped to a thickener, where the slurry is concentrated to about 30-percent solids and pumped to a vacuum filter. The solids cake on the filter is washed to recover soluble sodium. The filter cake, at 45 to 60 percent solids, is trucked to a one-acre disposal area.

Liquid filtrate and wash liquor from the filters are returned to the thickener. Sodium carbonate is fed to the thickener to make up any sodium losses. This is not intended as a softening step, since the Ca^{++} concentration is usually less than 100 ppm; but some $CaCO_3$ precipitation does occur. Clear liquor overflow from the thickener is fed to a holding tank for recycle to the absorber. The combined fresh feed and recycle streams have a pH from 6 to 11. Pertinent design data are presented in Table 3-26.

Distinguishing Features - This installation is characterized by:

- Operation in the concentrated mode using lime for regeneration.
- Firing with coal of 0.9- to 3.1-percent sulfur.
- Capability of several venturi/absorber configurations.

Table 3-26. CEA/ADL GULF POWER DEMONSTRATION UNIT: ORIGINAL DESIGN PARAMETERS[58]

Identification	Value
Boiler	Unit No. 1
Boiler capacity	
Nominal	40 MW
Maximum	47.5 MW
Fuel	Pulverized coal
Fuel characteristics	
HHV	28,400 J/g (12,200 Btu/lb)
Sulfur	3 percent
Ash	14 percent
Absorber inlet conditions	
Capacity	
Generating capacity	20 MW
Flow rate (standard)	25.5 m^3/s @ 21°C (54,000 scfm @ 70°F)
Flow rate (actual)	35.4 m^3/s @ 136°C (75,000 acfm @ 276°F)
Temperature	136°C (276°F)
Sulfur dioxide loading	1800 to 3800 ppm (dry)
Particulate loading	0.05 g/m^3 (0.02 gr/scf) dry
Oxygen	6.5 percent (max.)
SO_2 removal efficiency	90 percent
Maximum SO_2 removal rate	694 kg/hr (1530 lb/hr)
Maximum particulate loading	0.05 g/m^3 (0.02 gr/scf) dry
Power consumption, percentage of unit capacity	
Without venturi scrubber	1.0 to 1.5 percent
With venturi scrubber	2.5 to 3.0 percent

Operating History - The system started up February 3, 1975, and operated continuously through July 18, 1975, when it was shut down for repairs and modifications. The second period of operation was from September 16, 1975, through January 2, 1976.

The system exhibited excellent SO_2 removal capabilities, 90 percent and greater. Using the combined venturi/absorber configuration at a venturi liquor pH above 5.2, outlet SO_2 concentrations below 50 ppm were achieved; this corresponds to greater than 95-percent removal.[59] Raising the pH of the venturi liquor above 6.0 resulted in SO_2 removal efficiencies greater than 98 percent.[59] Figure 3-48 presents the relationship of scrubber bleed liquor pH to SO_2 outlet concentration.

Table 3-27 summarizes the operation of the Scholz facility with respect to various performance parameters.

The operability of the Scholz plant for the period February 1975 through June 1976 is presented in Table 3-28. The Scholz plant was designed to demonstrate the viability of the double alkali process technology for application on utility coal-fired boilers. As such, this prototype plant had less spare equipment than would be normal in full-scale applications. The operability of the system has been steadily improved; during the last four months of operation, it was 94 percent.

Operational Problems/Solutions - Problems occurring at the Scholz plant have been mainly mechanical and equipment-related.[61] These are summarized in Table 3-29, together with their solutions.

Summary - Operation of the Scholz prototype plant has demonstrated the viability of 90-percent SO_2 removal at high

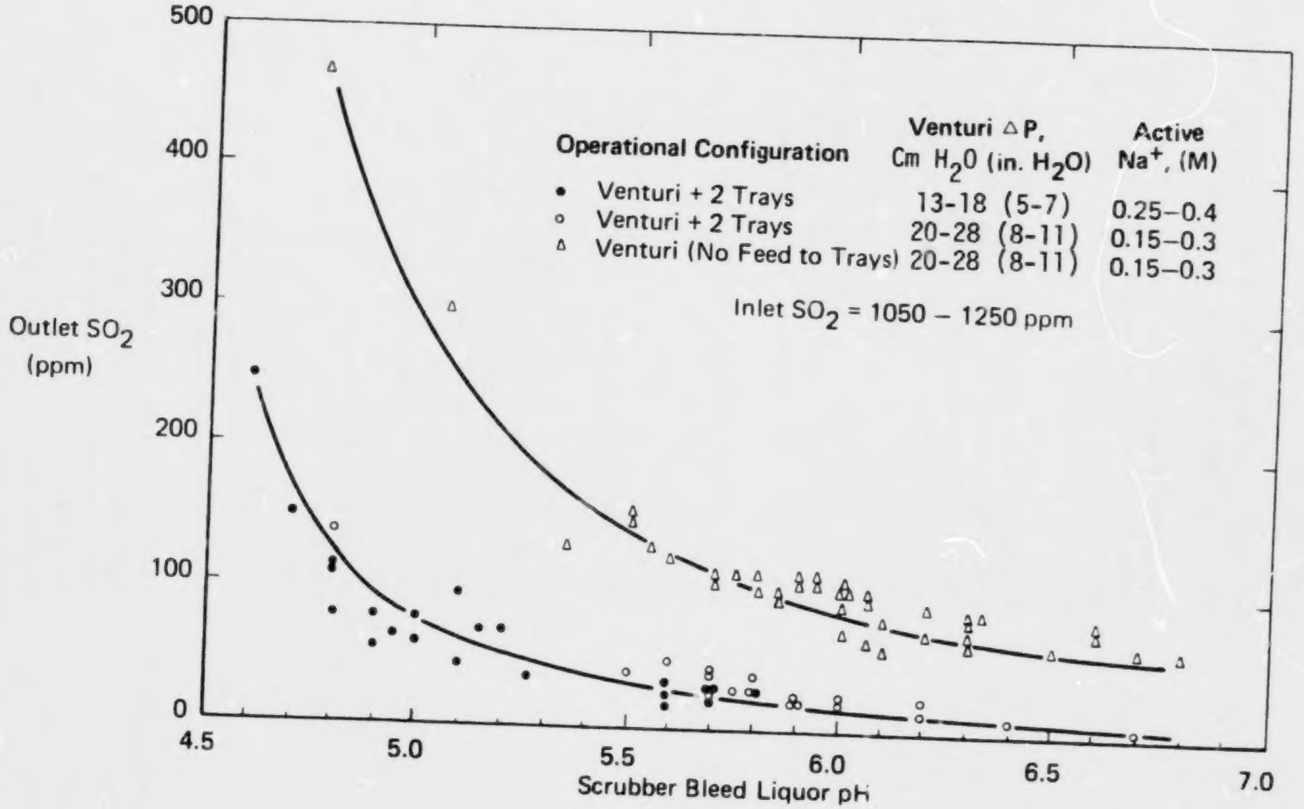

Figure 3-48. Outlet SO_2 concentration versus scrubber bleed pH.[63]

Table 3-27. CEA/ADL DOUBLE ALKALI PROTOTYPE SCRUBBER SYSTEM OPERATION RELATIVE TO MAJOR PERFORMANCE PARAMETERS[60]

SO_2 removal	The system generally operated at SO_2 removal efficiencies of about 95 percent and demonstrated the capability for removal of over 99 percent.
Particulate removal	As was intended in the design, there was no increase in loading.
Oxidation	Maximum oxidation rates of 25 to 30 percent in closed-loop operation were experienced without serious process upset.
Sodium makeup	Under these process conditions, approximately 0.03 mole Na_2CO_3 was required per mole of SO_2 removed as makeup for soluble solids losses in the washed cake. About an equal amount of sodium makeup was unaccounted for, probably because of leakage from the system and errors in soda ash makeup estimates.
Lime utilization	Lime utilization typically ranged from 95 percent to 100 percent. The overall lime stoichiometry was 0.95 to 1 mole $Ca(OH)_2$/mole of SO_2 removed from the flue gas.
Power consumption	This was 2.5 to 3 percent of the power generated by the utility boiler with the system operating at full fan and at full venturi and absorber liquid recirculation capacity. Correcting for the excess fan and pump capacity, the power consumed by the equipment actually required in the application (Tray tower at an L/G of 0.7 to 1.3 l/m^3 (5 to 10 gal/1,000 ft^3) is roughly 1 to 1.5 percent of the power generated at the design conditions.
Filter cake properties	Waste Solids Properties - The system produced a washed filter cake generally containing at least 50 percent insoluble solids. The waste material, on the average, contained 3 to 5 percent soluble solids (dry cake basis).

Table 3-28. CEA/ADL DOUBLE ALKALI PROTOTYPE SCRUBBER PERFORMANCE HISTORY: OPERATION AND VIABILITY PARAMETERS

Period	Total period, hr	Boiler operation, hr	FGD operation, hr	FGD[a] operability, percent	FGD[b] utilization, percent
Feb. 75	672	459	454	98.9	67.6
Mar. 75	744	507	485	95.7	65.2
Apr. 75	720	604	336	55.6	45.2
May 75	744	598	375	63.2	50.4
Jun. 75	720	720	720	100.0	100.0
Jul. 75	744	683	221	32.4	29.7
Aug. 75	744	744	0	0	0
Sep. 75	720	577	254	44.0	35.3
Oct. 75	744	559	559	100.0	75.1
Nov. 75	720	620	560	90.3	77.8
Dec. 75	744	732	732	100.0	98.4
Jan. 76	744	0	0	0	0
Feb. 76	696	0	0	0	0
Mar. 76	744	480	445	92.7	59.8
Apr. 76	720	642	616	95.9	85.6
May 76	744	735	651	88.6	87.5
Jun. 76	720	656	641	97.7	89.0

[a] FGD operability: The number of hours the FGD system was in operation divided by the number of hours the boiler was in operation, expressed as a percentage.

[b] FGD utilization: The number of hours the FGD system was in operation divided by the number of hours in the period, expressed as a percentage.

Table 3-29. SUMMARY OF THE PROBLEMS AND SOLUTIONS IN OPERATION OF CEA/ADL DOUBLE ALKALI PROTOTYPE SCRUBBER

Unit operation area	Problem	Solution
Scrubber-absorber module	Failure of rubber linings in bleed control valves	Replaced with stainless steel valves
Regeneration reactor	Plugging of dry lime feed chute to first stage of multistage reactor	Installation of feed vibrator
	Solids deposition in first stage of multistage reactor	Design and installation of a new first-stage unit
	Broken agitator blade in second stage of multistage reactor	Part replacement
	Erosion and plugging of flow-through pH meters	Replaced with immersion type meter with sonic cleaner
Thickener	Plugging of thickener underflow lines	Installation of flexible lines and addition of back-flushing procedures
	Deterioration of thickener	Patch/replace lining section
Vacuum filter	Erosion and cracking of fiberglass scraper and rocker arm	Replaced with stainless steel parts
	Erosion of plastic bridge valve	Part replacement
	Holes in filter cloth	Repaired holes
	Loss of vacuum due to cracks in trunnion tubes	Patched cracks
	Insufficient agitation	New agitation system
Sodium carbonate addition	Plugging in dry feeder gate	Replaced feed control unit
	Insufficient reliability of sodium makeup feed solution	Revised monitoring system and employed manual techniques

Table 3-29 (Cont'd). SUMMARY OF THE PROBLEMS AND SOLUTIONS IN OPERATION OF CEA/ADL DOUBLE ALKALI PROTOTYPE SCRUBBER

Unit operation area	Problem	Solution
Reheater	Deterioration of refractory in burner chamber	Replaced refractory material
Fans	Fly ash buildup on fan blade	Clean and rebalance fan
	Deterioration of thrust bearing	Part replacement
Pumps	Leakage of scrubber recycle pumps through pump seals and seals and piping	Replaced packing and flange gaskets
	Broken shaft on reactor pump	Part replacement
	Failure to reactor pump isolation valve	Overhauled
	Freezing of pump seal water rotameters	Replaced parts and modified operating procedures
Tanks	Cracks and pinholes in the lining of the absorber recycle tank	Repaired
	Plugging of pH meter on recycle tanks	Repaired and relocated sampling lines
	Liner failure in the thickener hold tank	Replaced
	Insufficient reliability of level transmitter in the thickener hold tank	Replaced
Piping	Circuitry failure in heat tracing	Replaced
Stack	Failure of stack lining	Repair/replace

availabilities on high- and low-sulfur coal applications. The success of the prototype facility has enabled a full-scale system to be installed on a utility boiler (Louisville Gas and Electric's Cane Run 6).

3.3.3 Description of Japanese Double Alkali FGD Installations

There are 19 plants in Japan on which sodium-based double alkali systems are operating. The plants differ from U.S. applications in that by-product gypsum is produced rather than disposable sludge. Table 3-30 lists the systems and their applications. Of these installations, 13 are industrial-size boilers, five are utility boilers, and one is a sulfuric acid plant. All boilers are fired with oil, but the principle of operation and equipment used is virtually identical to operation on coal-fired units.

Following is a brief description of the systems, as offered by the suppliers, and a summary of system performance in Japan.

3.3.3.1 Kawasaki/Kureha[63] - Kureha Chemical Industry Co., Ltd., and Kawasaki Heavy Industries, Ltd., have jointly developed a commercial double alkali process utilizing concentrated sodium solution and limestone regeneration. The system is currently installed on five oil-fired utility boilers.

System Description - A typical schematic diagram of the system as installed at Tohoku Electric is shown in Figure 3-49. The system consists of an SO_2 absorption section, a limestone regeneration section, a sulfate removal section, and a gypsum production section.

There is a precooler section at the entrance to the scrubber, in which the hot incoming gas from the boiler is

Table 3-30. SODIUM-BASED DOUBLE ALKALI PROCESS INSTALLATIONS IN JAPAN[62]

Process developer	Absorbent, precipitant	User	Plant site	Capacity, 1,000 Nm^3/hr	Source of gas	Inlet SO_2, ppm	Year of completion
Showa Denko	Na_2SO_3, $CaCO_3$	Showa Denko	Chiba	500	Industrial boiler	1,500	1973
		Kanegafuchi Chem.	Takasago	300	Industrial boiler	1,500	1974
		Showa Pet. Chem.	Kawaski	200	Industrial boiler	1,400	1974
Showa Denko-Ebara	Na_2SO_3, $CaCO_3$	Nippon Mining	Saganoseki	120	H_2SO_4 plant		1973
		Yokahama Rubber	Hiratsuka	105	Industrial boiler		1974
		Misshin Oil	Isogo	100	Industrial boiler		1974
		Poly Plastics	Fuji	212	Industrial boiler		1974
		Ajinomoto	Yokkaichi	82	Industrial boiler		1974
		Kyowa Pet. Chem.	Yokkaichi	150	Industrial boiler		1974
		Japan Food	Yokkaichi	100	Industrial boiler		1975
		Yokohama Rubber	Mie	100	Industrial boiler		1975
		Asia Oil	Yokohama	243	Industrial boiler		1975
Kureha-Kawasaki	Na_2SO_3, $CaCO_3$	Tohoku Electric	Shinsendai	420	Utility boiler	420	1974
		Shikoku Electri	Sakaide	1,260	Utility boiler	1,500	1975
		Shikoku Electric	Anan	1,260	Utility boiler	1,500	1975
		Myushu Electric	Buzen	730	Utility boiler	1,500	1977
		Tohoku Electric	Akita	1,050	Utility boiler	1,500	1977
Tsukishima	Na_2SO_3, CaO	Kinuura Utility	Nagoya	185	Industrial boiler		1974
		Daishowa Paper	Fuji	264	Industrial boiler		1975

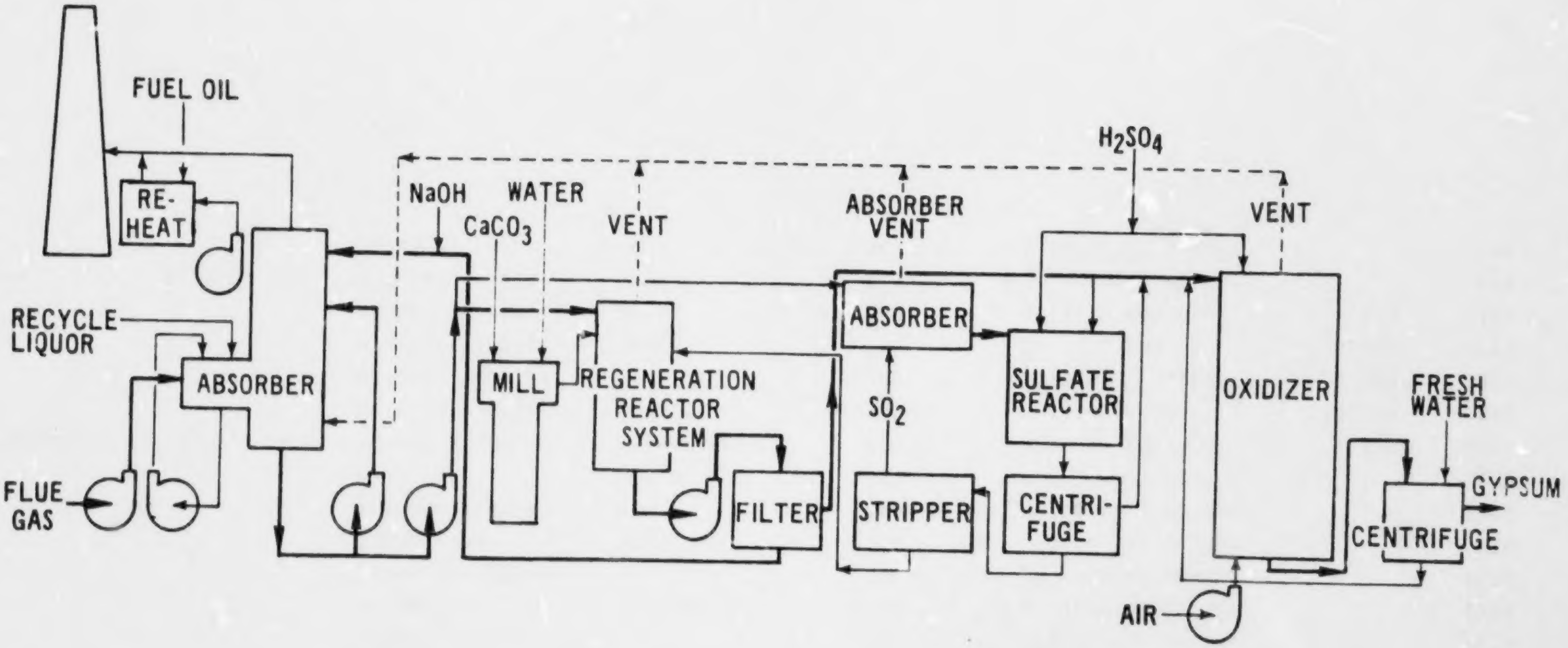

Figure 3-49. Kureha-Kawasaki double alkali process at Tohoku Electric's Shinsendai station.[64]

adiabatically cooled with recycle liquor. The scrubber is of the grid-packed type, consisting of two separate stages of packing; gas flows upward through the packing. Scrubber recycle liquor and effluent from the top stage are fed to the top of the bottom stage which is operated at an L/G ratio of about 1.6 l/m^3 (12 gal/1000 ft^3). Regenerated liquor alone is fed to the top stage, thus maximizing the pH in this stage and achieving good countercurrent mass transfer; this results in high SO_2 removal efficiency. Clean flue gas leaving the absorber is reheated to 141°C (285°F) by the direct firing of oil.

The regeneration reaction consists of neutralization of $NaHSO_3$ with finely ground limestone to regenerate sodium hydroxide. The limestone is ground to 10 μm in a wet, vertical tower mill. Scrubber effluent liquor bleed taken from the bottom-stage recycle stream is filtered to remove oil soot before it is pumped to the regeneration reactors to be reacted with limestone. The reactor system consists of five closed reactors in series, operated at about 90°C to promote the reaction between limestone and $NaHSO_3$ in the scrubber bleed. The solid product produced in the reactors is separated from the mother liquor and washed in a vacuum filter. The liquor is recycled to the scrubber system, and the solids, mostly $CaSO_3 \cdot 1/2\ H_2O$, are fed to the oxidizer system and to the sulfate removal section.

This system is designed to remove sulfate formed in the scrubber as gypsum through reaction with sulfuric acid and calcium sulfite. The process is fed with a bleed stream of scrubber effluent, a portion of the calcium sulfite cake (washed) from the filter and sulfuric acid. In order to maximize sulfuric acid efficiency, a small absorber and stripper are incorporated in the sulfate treatment loop. The incoming scrubber bleed is acidified in the absorber with SO_2; this SO_2 is supplied by the stripper, which converts $NaHSO_3$ to Na_2SO_3, stripping off SO_2 by means of thermal decomposition according to the following reaction:

$$2NaHSO_3 \xrightarrow{heat} Na_2SO_3 + SO_2 + H_2O \quad (19)$$

Feed liquor to the stripper is produced by the acidic mother liquor from the sulfuric acid reactor in which the sulfate removal occurs. Stripped liquor is recycled to the regeneration reactor system.

Feed to the sulfuric acid reactor is thus already at a low pH, and therefore requires less sulfuric acid addition to achieve a given conversion of Na_2SO_4 to gypsum.

Effluent slurry from the sulfuric acid reaction is centrifuged to separate the gypsum produced from acidic mother liquor. The separated solids are transferred to the oxidizer system, together with the calcium sulfite solids produced in the regeneration reaction. The centrifugate, as previously stated, is recycled to the stripper to allow recovery of acid values.

The oxidizer developed by Kawaski/Kureha is considered to be proprietary. Oxidation air is supplied at 5 psig. Repulped calcium sulfite slurry from regeneration, and gypsum from the sulfate-removal step are pumped to the oxidizer. Sulfuric acid is added to adjust pH in the reaction to the right conditions for the oxidation reaction. Calcium sulfite is oxidized to gypsum by reaction with atmospheric oxygen according to the overall reaction:

$$CaSO_3 \cdot 1/2H_2O + 1/2O_2 + 3/2H_2O \rightarrow CaSO_4 \cdot 2H_2O \quad (20)$$

The reaction mechanism may be acidic dissolution of $CaSO_3 \cdot 1/2H_2O$, followed by oxidation of the bisulfite to sulfate which regenerates the acid value to dissolve additional $CaSO_3 \cdot 1/2H_2O$. Gypsum product is separated from oxidizer effluent and washed in centrifuges. Spent off-gas from the oxidizer is recycled through the scrubber to remove SO_2 that was released by the acidic oxidation reaction.

Typical conditions and performance data for the Tohoku Electric FGD system are given below:[65]

Flue Gas

Source: oil fired boiler
Fuel: 1.2 to 1.5 percent sulfur oil
Oxygen content: 3 percent
Flow rate: 420,000 m^3/hr (247,000 scfm)
SO_2 inlet concentration: 600 to 800 ppm
Particulate inlet: 2 to 10 mg/m^3 (0.0009 to 0.0044 gr/scf)
Reheat: direct fire with 1.2 percent sulfur oil to 140°C
(285°F)

Absorber System

Absorber type: grid-packed, two packing stages
SO_2 removal: 98 percent
SO_2 exit concentration: <10 ppm
L/G ratio: 1.6 l/m^3 (12 gal/1000 ft^3, bottom stage
Scrubber pressure drop: 200 mm H_2O (7.9 in. H_2O
Scrubber liquor pH: 6.9 to 7.6 inlet to top stage
6.3 to 6.5 bottom stage recycle
Recycle liquor composition: 5 to 10 percent Na_2SO_4
5 to 8 percent $NaHSO_4$
6 to 9 percent Na_2SO_3
Oxidation rate: 5 percent

Regeneration

Limestone stoichiometry: 100 percent of SO_2 inlet
Sodium makeup rate: 0.018 moles Na/mole SO_2 absorbed
Reactor system residence time: 3 hr (total, five reactors)
Reactor percentage solids: 5 percent (essentially all
$CaSO_3 \cdot 1/2H_2O$)
Reactor filter cake solids: 40 percent
Regenerated liquor: 6.9 to 7.6 pH
20 percent total dissolved solids
<10 ppm Ca
Limestone particle size: 10μm

Solids Production (Gypsum)

Water content: 5 percent
Sodium content: <300 ppm on dry bases
Particle size: 30 x 200μm

The system is said to be a closed-loop system, in that there is no intentional water purge; however, based on the rate of water addition to the system, and a comparison of the rate of sodium makeup with the sodium content of the product, it appears that there is some sodium loss from the system other than the amount lost in the product gypsum. The overall system is fairly complex when compared with some of the systems under development in the United States. The reasons for the additional complexity appear to be related to the following: a) the system produces a relatively pure, salable product, b) it is designed to achieve extremely high SO_2 and particulate removal, and c) limestone rather than lime is used as the source of calcium. Apparently, considerable effort has been devoted to minimizing the amount of reagent used; i.e., sulfuric acid, limestone, and sodium.

The gypsum product has been used in cement manufacture; however, it also meets the more stringent specifications for wallboard production.

The very low oxidation (5%) in the absorption system is attributed by Kawasaki/Kureha to "proprietary know-how."

3.3.3.2 Showa Denko KK/Ebara[66] - Showa Denko KK and Ebara Manufacturing Co., Ltd., Japan, jointly developed a concentrated double alkali process using limestone as the calcium source, and producing a salable gypsum by-product. The system is currently operating on 11 industrial oil-fired boilers and a sulfuric acid plant.

Figure 3-50 is a schematic representation of the Showa Denko/Ebara process. Basically it is similar to the Kawasaki/Kureha system. The full-scale unit (150 MW) uses four inverted "vertical cone" scrubbers (similar to the Zurn type). Liquor is entrained at the bottom of a conical draft tube (truncated apex) by the entering flue gas and the mixture passes up the tube. At the top of the tube (conical base) there is a disengaging section, where liquor separates and falls back to the bottom of the scrubber.

In the reactor system, limestone is reacted with spent scrubber effluent to neutralize the $NaHSO_3$, producing Na_2SO_3 for recycle to the absorber, and $CaSO_3 \cdot 1/2H_2O$, which is sent to an air oxidizer to be converted to gypsum by air oxidation (equation 28). A slipstream of spent scrubber liquor is taken to a sulfate treatment section, which converts Na_2SO_4 to gypsum and $NaHSO_3$ by reacting it with sulfuric acid and calcium sulfite. The gypsum, which contains some unconverted calcium sulfite, is combined with the calcium sulfite produced in the limestone neutralization reaction and pumped to the air oxidizer system. Sulfuric acid is added to adjust the pH in the oxidizer system so that optimum conditions can be achieved for calcium sulfite oxidation.

The system operates in an open loop, since the amount of water required to wash the gypsum product to below 300-ppm sodium content is greater than water losses through evaporation, water of hydration, and moisture in the final product. The sodium content specification for wallboard manufacture is less than 300 ppm. The corresponding specification for gypsum used in cement manufacture is 900 ppm.

General conditions and performance data for the full-scale system are as follows:

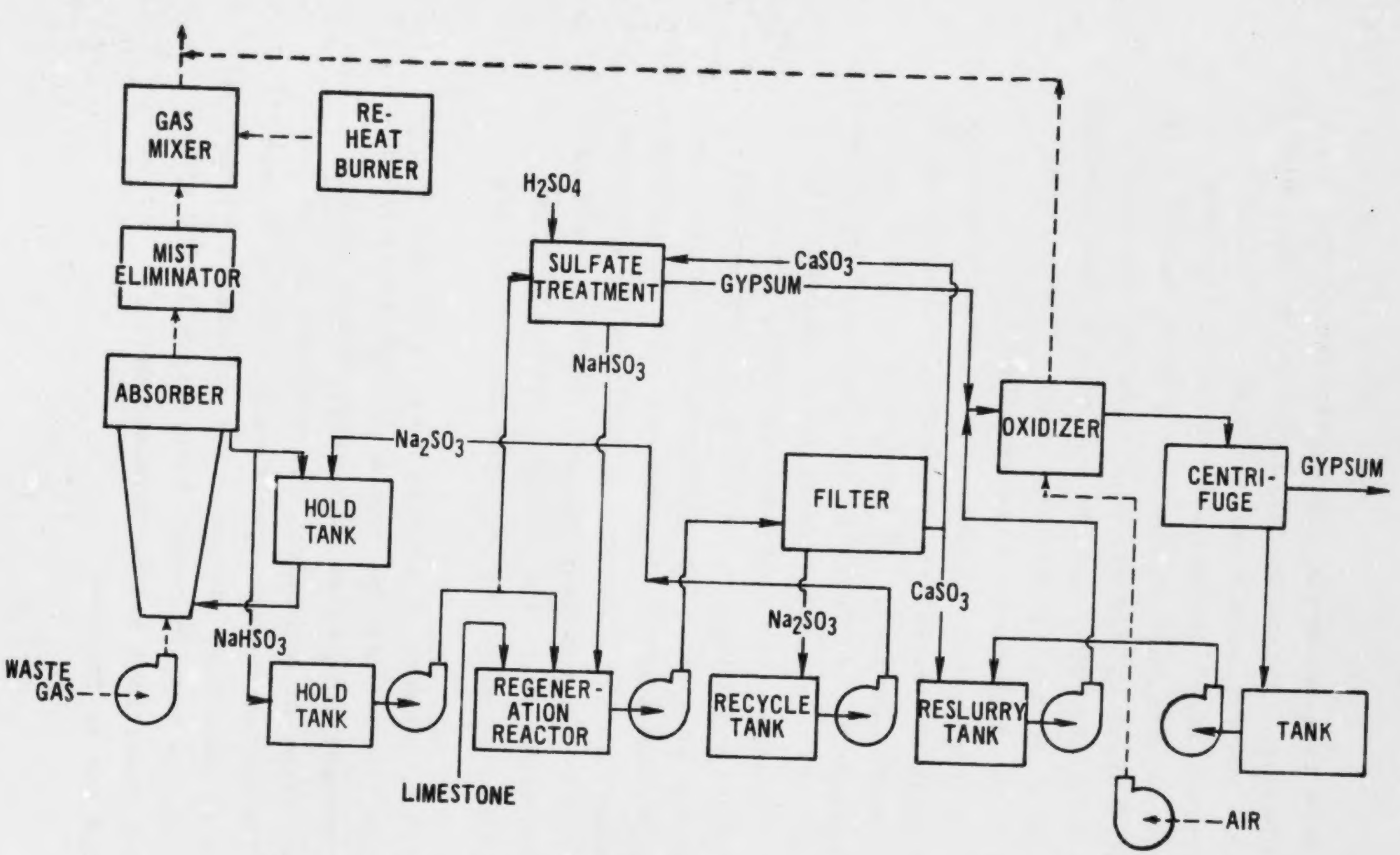

Figure 3-50. Showa Denko sodium-limestone process.[67]

Flue Gas

Source: oil-fired 150-MW boiler
Fuel: 2.5 to 3.0 percent sulfur
Particulate inlet concentration: 1.8 g/m^3 (0.8 gr/scf)
SO_2 inlet concentration: 1200 to 1500 ppm
Reheat: direct-fired with 1 percent sulfur oil to 141 to 149°C (285 to 300°F)

Absorber System

Absorber type: four vertical cone scrubbers
SO_2 exit concentration: 60 to 100 ppm after reheat with 1 percent sulfur oil
Particulate exit concentration: 0.057 to 0.80 g/m^3 (0.025 to 0.035 gr/scf)
L/G ratio: 1 l/m^3 (7.5 gal/1000 ft^3)
Scrubber liquor pH: 6.3 exit
Oxidation rate: <10 percent

Regeneration Chemical Usage

Limestone utilization: 96 percent
Sulfuric acid usage: 0.2 moles H_2SO_4/mole SO_2 absorbed
Sodium usage: 0.10 to 0.12 moles of NaOH/mole SO_2 absorbed

Oxidizer

Suspended solids content: 15 percent

Gypsum By-Product

Na content: <300 ppm
Water content: 7 to 8 percent

3.3.3.3 Performance of Double Alkali Systems in Japan - The processes offered by Kureha-Kawasaki and Showa-Denko are very effective for removal of SO_2. At the Chiba Plant, the Showa-Denko process has demonstrated 95-percent SO_2 removal on a flue gas stream containing 1200 to 1500 ppm SO_2.[68] At the Shinsendai plant, the Kureha-Kawasaki process has demonstrated SO_2 removals of over 97 percent on a flue gas stream containing 600 to 800 ppm.[69] No detailed information is available on reliability, but it is reported that it approaches 100 percent.[70]

3.3.4 Engineering Design Parameters

Sulfur dioxide efficiencies of over 90 percent can be achieved with sodium-based double alkali FGD system on coal-fired boilers firing high- or low-sulfur coal. The key operating parameters may vary depending on the mode of operation chosen (i.e. dilute vs. concentrated). In the United States, only two basic scrubbing strategies have been investigated: dilute active sodium with lime regeneration, and concentrated active sodium with lime regeneration. Of these two, the concentrated mode has received the major emphasis in research and development on coal-fired boiler flue gas applications and is the mode selected for full-scale operation at the three utility boiler applications described in Section 3.3.2.

Because of the high solubility of sodium alkalis and their sulfur-containing reaction products, problems associated with plugging, scaling, and limited removal capability are minimized. In regard to SO_2-removal capability and scale formation, the scrubbing solution pH range provides the key to optimal system efficiency. A pH control value of 6.5 in the scrubbing effluent allows the achievement of maximum SO_2 removal efficiency without the occurrence of chemical-related problems. If the scrubbing solution pH value exceeds 7, CO_2 absorption would become significant and could lead to calcium carbonate scaling. A scrubbing solution pH level below 6 is likewise avoided, because the vapor pressure of sulfur dioxide increases dramatically for concentrated active alkali systems and can lead to equilibrium-limited scrubbing conditions where outlet concentrations at or below 200 ppm are desired. This chemical phenomenon is demonstrated in Figure 3-51, where the sulfur dioxide vapor pressure is plotted as a function of pH for a solution temperature of 54°C (130°F), a typical saturation temperature for a boiler flue gas stream. For a concentrated active alkali system, the vapor pressure of SO_2 as a function of pH is significant. The SO_2 vapor pressure at a pH of 6 is approximately 100 ppm and increases rapidly as the pH drops.

For these reasons, concentrated systems tend to operate in the pH range of 6 to 7, and preferably at a set point of 6.5, where the sulfite-bisulfite system is highly buffered. The highly buffered system is preferred, because the scrubbing solution is able to adapt to rapid changes in flue gas inlet conditions. This is in contrast to scrubbing systems using a pH of 6 or less; these absorbents can be very sensitive to rapid changes in flue gas inlet concentrations.

Thus, based upon the low SO_2 back pressure of the reacted species in solution at the optimum pH control value of 6.5, SO_2 concentration levels in the scrubber discharge can be maintained as low as 10 to 20 ppm.

The principal area of concern in regard to the chemical characteristics of double alkali processes is their tendency to form inactive species through reaction with oxygen or chlorides. Oxidation in a double alkali system refers to the conversion of active sulfite and bisulfite species to inactive sulfate. This can occur either in the gas ($SO_2 + 1/2\ O_2 \rightarrow SO_3$) or aqueous phase ($SO_3^= + 1/2\ O_2 \rightarrow SO_4^=$; $HSO_3^- + 1/2\ O_2 \rightarrow SO_4^= + H^+$).

The formation of inactive sulfate in the scrubbing solution can have a deleterious effect on process chemistry

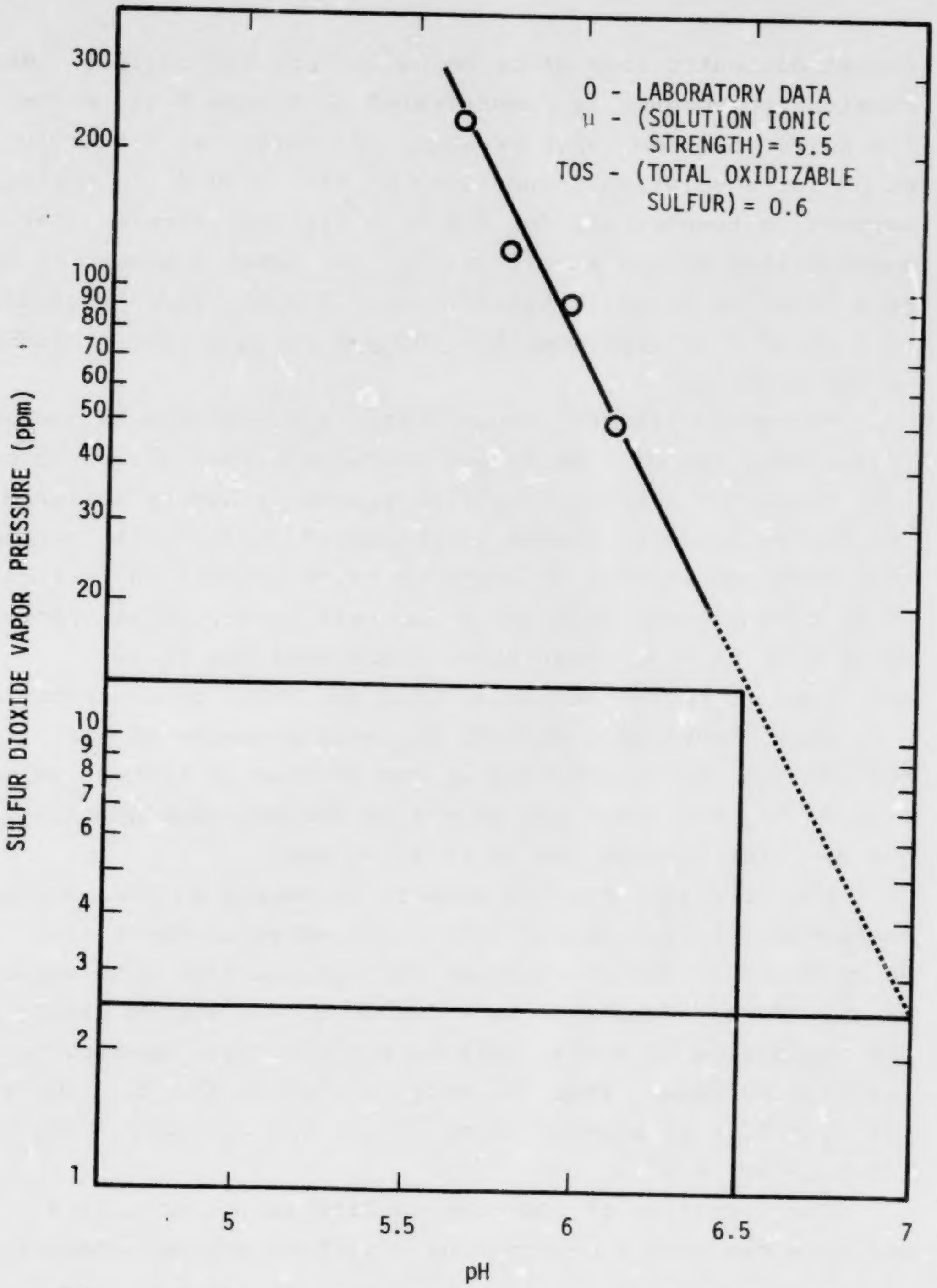

Figure 3-51. Concentrated active alkali system: sulfur dioxide vapor pressure vs. pH @ 54°C (130°F).[71]

because of a) its inability to react with SO_2, b) a reduced ability to be regenerated into an active alkali form by reaction with lime, and c) the potential to form gypsum scale under certain conditions. Oxidation may occur in any segment of the scrubber plant; it is generally considered to be a function of the rate of dissolution of oxygen; scrubbing solution pH; trace contaminants (catalytic effects of collected fly ash; lime and soda ash impurities); solution ionic strength; absorber design; SO_2 concentration of the flue gas; O_2 concentration of the flue gas; and flue gas temperature. Evidence indicates, however, that approximately 90 percent of total system oxidation occurs in the scrubber circuit.[72] Sulfates along with sulfites are removed from the spent scrubbing solution by reaction with lime in the regeneration/precipitation reactor. In concentrated mode systems, the unsaturated calcium sulfate mode of operation is maintained by coprecipitation of calcium sulfite and sulfate from slurry liquid that is saturated in calcium sulfite and unsaturated in calcium sulfate and favored by low sulfite-to-sulfate oxidation tendencies. Calcium sulfate can be coprecipitated with the principal sulfite precipitate up to a sulfate/sulfite mole ratio of approximately 0.20. Thus, unsaturated conditions can be maintained only at relatively low sulfite-oxidation rates, achievable in high SO_2/O_2 ratio emission sources. The minimum inlet gas conditions that must be met in order to operate a concentrated mode system reliably and economically, are about 400 to 500 ppm SO_2 and 5 percent O_2. For flue gas applications that equal or fall below these inlet gas requirements (such as predominate in some industrial boilers, bark boilers, and paper mills), dilute mode double alkali systems are technically feasible, owing to their ability to withstand high oxidation rates (up to 100 percent oxidation).

Inactive sodium chloride may also be formed in the system as a result of absorption of hydrogen chloride by the scrubbing liquor and the subsequent formation of hydrochloric acid (HCl) and NaCl. Where filter cake washing is used to recover soluble sodium, higher NaCl concentrations can build up in the liquor recycle and reduce SO_2 removal efficiency, brought about by loss of active sodium. Purge streams or prior CHl removal must be practiced to minimize this effect.

The double alkali system must be designed to limit dissolution of oxygen, so that oxidation of sulfite/bisulfite to sulfate can be minimized. The chloride level in the scrubber circuit must also be controlled by purging, in order to hold down sodium losses via soluble sodium chloride; to limit corrosion attack on scrubber internals; and to permit operation in a mode that is sulfate-unsaturated. Sophisticated solids dewatering is also necessary to prevent soluble sodium losses, thereby limiting potential environmental problems associated with waste disposal and permitting closed-loop operation. Finally, instrumentation for pH control of the absorber and regeneration systems must be reliable to insure adequate SO_2 removal in a scale-free environment.

As stated previously, 90 percent of the oxidation of sulfites to sulfates occurs in the scrubber circuit. Since oxidation is a strong function of the rate of oxygen absorption, the number of contact stages in the scrubber module should be limited to two (e.g., the FMC, CEA/ADL and Envirotech full-scale electric utility applications all incorporate two-stage modules).[73] In addition, the regeneration reactor must also employ an optimum residence time to minimize the oxidation occurring in the reactor.[74]

The dry removal of fly ash in a high-efficiency ESP upstream from the scrubber plant also helps to limit sulfite oxidation caused by catalyzing metal compounds.

The presence of chloride in the coal at or above 0.04 percent by weight can result in increased scaling in the scrubber loop, attributable to sodium sulfate insolubility, which itself is caused by high sodium chloride concentrations (1.5 to 2.0 moles).[75] In addition, all hydrogen chloride absorbed in the scrubber circuit ties up the sodium as dissolved sodium chloride, which will be completely lost as inactive sodium. This increases the potential for pollution problems in surrounding ground and surface waters, and also results in a higher sodium consumption. These considerations, plus the possibility of chloride corrosion, may force the inclusion of a separate water loop scrubber before the SO_2 absorber, in order to remove most of the chlorides from the flue gas.

Solids separation by means of a rotary drum filter, belt filter, extraction filter, or centrifuge, in addition to one or more thickeners, is necessary in order to operate a closed-loop system, to minimize potential water pollution problems, and to reduce soda ash consumption.

As stated previously, pH control in the SO_2 absorption step determines the amount of SO_2 removed and precludes the formation of carbonate scale. Also, pH control in the regeneration reactor determines the composition of the regenerated solution and precipitated solids. A pH of 8.5 is the titrametric end-point for sodium bisulfite. It is possible to precipitate additional sulfur and to regenerate active sodium by controlling at more alkaline pH levels (11 through 12.5). The pH is more difficult to control in the more alkaline regions, however, because of poorer lime system responsiveness to solution pH values, resulting in higher stoichiometric lime requirements, lower lime utilization values, poorer filter cake quality, and a potential for calcium ion supersaturation in the scrubber circuit. The pH control loops must be designed and situated to respond reliably and quickly if optimum system operations are to be ensured.

In summary, a double alkali system designed to effect 90-percent SO_2 removal or greater on a high- or low-sulfur coal-fired utility boiler would have the following design parameters:

- Operation in the concentrated alkali mode (e.g. >0.15M active alkali concentration in the scrubber feed).
- Use of a venturi scrubber with a separate water loop for control of particulates and chlorides for high-chloride coal applications (i.e. >0.04% Cl by weight in the coal).
- Use of a two-stage tray or packed tower absorber with an L/G ratio of from 1.3 to 2.7 l/m^3 (10 to 20 gal/1000 acfm), a scrubbing liquor pH of 6.0 to 7.0, and a pressure drop of 15 to 30 cm (6 to 12 in.) H_2O.
- Regeneration of the active sodium through addition of lime or limestone to a reactor tank having a residence time of 10 min for lime and more than 30 min for limestone. Utilization rates of 90 percent or greater should be obtained for lime and 75 to 85 percent for limestone.
- Sodium makeup by addition of sodium carbonate with its associated softening effect. Sodium makeup should be equivalent to 0.05 moles of sodium carbonate per mole of SO_2 removed.

- The solids removal section should consist of clarifiers and vacuum filters to ensure high solids content in the final waste cake. Water washing of the waste cake should be used to recover soluble sodium present in the cake.
- Operation should be in a closed-loop mode with respect to the water balance by use of recycle water.

3.3.5 Process Operability

Although no full-scale double alkali FGD systems are in operation on coal-fired boilers, it is possible to predict operability of the systems, based on experience with smaller units on coal-fired boilers. As shown in Section 3.3.2, the operability of double alkali FGD systems on coal-fired utility boilers and prototype utility installations has been improved steadily; it is now 90 percent and above. Most operability problems were due to design-related equipment shortcomings in these prototype installations. It should be pointed out that most installations did not have spare equipment; this would, however, be included in full-scale utility systems.[76]

Equipment-related problems were solved at each of the double alkali installations and the experience gained will benefit later installations. The vendors of double alkali systems have developed confidence in their reliability: as evidenced by guarantees of 90-percent availability for the first year of operation and 100 percent for the life of the plant (based on a boiler operating rate of 70 percent for some of the new, full-scale utility applications). The systems are all guaranteed to achieve 85- to 95-percent SO_2 removal efficiency on high-sulfur coal applications. No new low-sulfur coal applications are planned, but similar guarantees would be expected for such systems.

Corrosion, erosion, and scaling problems have not been important factors at double alkali FGD installations. Full-scale versions of these systems are not expected to experience these problems either.

The double alkali system has demonstrated the ability to perform well under fluctuating SO_2 inlet concentrations. At the Scholz plant, the design inlet SO_2 concentration was 1800 ppm. At inlet concentrations varying from 800 to 1700 ppm, removal efficiencies were above 90 percent.[77]

Sludge produced in the double alkali FGD process must be disposed of safely. FMC Corporation has studied the properties of the sludge produced by their double alkali process. The study has revealed that the waste cake was highly impermeable; the only potential problem was the leaching of soluble salts contained in the cake.[78] This can be overcome by proper landfill management and/or the use of a liner in the disposal area. The waste cake exhibits good mechanical properties, making it unnecessary to use fixation to obtain mechanical stability.

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3.4 MAGNESIUM OXIDE SYSTEMS

This scrubbing system uses a slurry of magnesium oxide in water to absorb SO_2. The reaction product, magnesium sulfite, is then dried, and the magnesium oxide is regenerated and recirculated. Sulfur dioxide is recovered for conversion to sulfuric acid or reduction to sulfur.

3.4.1 Process Description

In this system, fly ash and other particulates are removed from the flue gas stream prior to the scrubbing operation, either by electrostatic precipitators (ESP's) or by primary (wet) scrubbers. The fly ash is removed to reduce solids loading in the absorber and to reduce impurities in the regenerated magnesium oxide.

The flue gas then enters an absorber for SO_x removal as shown in Figure 3-52. In this system, reactive magnesium oxide is slaked (reacted) with water to form magnesium hydroxide [$Mg(OH)_2$] slurry, which becomes the absorbent for the removal of SO_x. Sulfur dioxide and the small amount of SO_3 in exhaust gases from the boiler react with the magnesium hydroxide slurry to form magnesium sulfite ($MgSO_3$) and magnesium sulfate ($MgSO_4$) respectively. Some of the SO_2 reacts with the magnesium sulfite in the presence of water to form magnesium bisulfite [$Mg(HSO_3)_2$]. An excess of magnesium oxide (2.5 - 5%) is maintained in the slurry beyond that required to react with all the SO_x in the flue gas. The excess magnesium oxide or hydroxide reacts with the magnesium bisulfite, forming more magnesium sulfite. Additional magnesium sulfate is formed by the oxidation of magnesium sulfite.

Sulfur oxides absorption is represented by the following reactions:[1,2]

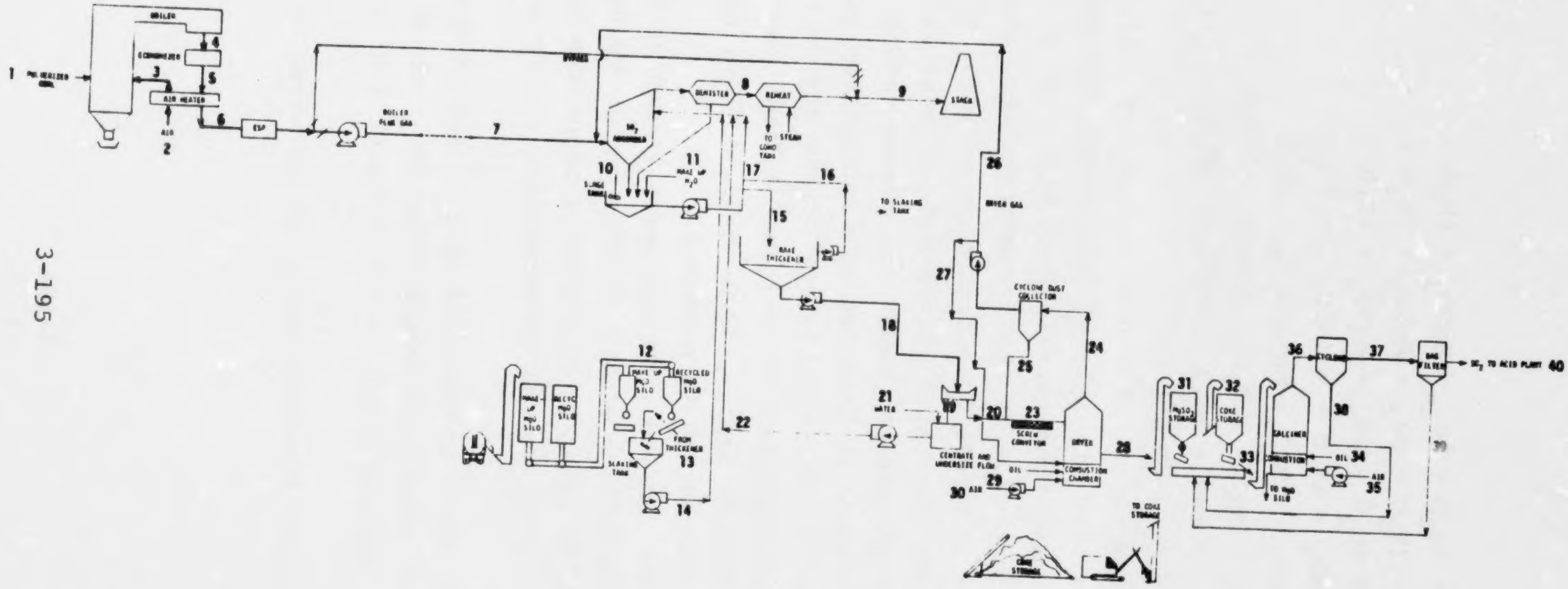

Figure 3-52. Magnesium oxide slurry FGD system.

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$Mg(OH)_2 + 5H_2O + SO_2 \rightarrow MgSO_3 \cdot 6H_2O \downarrow$ (1)

$Mg(OH)_2 + 2H_2O + SO_2 \rightarrow MgSO_3 \cdot 3H_2O \downarrow$ (2)

$Mg(OH)_2 + 6H_2O + SO_3 \rightarrow MgSO_4 \cdot 7H_2O \downarrow$ (3)

$SO_2 + MgSO_3 \cdot 6H_2O \rightarrow Mg(HSO_3)_2 + 5H_2O$ (4)

$SO_2 + MgSO_3 \cdot 3H_2O \rightarrow Mg(HSO_3)_2 + 2H_2O$ (5)

$Mg(HSO_3)_2 + MgO + 11H_2O \rightarrow 2\ MgSO_3 \cdot 6H_2O \downarrow$ (6)

$Mg(HSO_3)_2 + MgO + 5H_2O \rightarrow 2MgSO_3 \cdot 3H_2O$ (7)

$2\ MgSO_3 + O_2 \rightarrow 2MgSO_4$ (8)

The aqueous slurry absorbent (scrubbing medium) contains both hydrated crystals of MgO, $MgSO_3$, and $MgSO_4$, as well as a solution that is saturated with each of these components. A portion of the scrubber stream, theoretically the equivalent (expressed as magnesium sulfite) of the sulfur oxide content being introduced, is constantly diverted to a clarifier/thickener for concentration, and the concentrated slurry is then fed to a continuous centrifuge. Alternatively, the scrubber slurry stream may be fed directly to the continuous centrifuge. A "wet cake" containing crystals of $MgSO_3 \cdot 6H_2O$, $MgSO_3 \cdot 3H_2O$, $MgSO_4 \cdot 7H_2O$, and unreacted MgO is produced.

The liquor removed from the crystals may be returned to the main recirculating slurry stream, or may be used to slake/react with fresh or regenerated magnesium oxide (MgO), which is then added to the recirculating slurry stream as makeup.

The "wet cake" is conveyed to a direct-contact dryer where free and chemically bound water are removed. Dryer temperatures typically are in the range of 176-232°C (350-450°F).

The chemical reactions occur in the dryer as follows:

$MgSO_3 \cdot 6H_2O \overset{\Delta}{\rightarrow} MgSO_3 \cdot 3H_2O + 3H_2O \uparrow$ (9)

$MgSO_3 \cdot 3H_2O \overset{\Delta}{\rightarrow} MgSO_3 + 3H_2O \uparrow$ (10)

$MgSO_3 \cdot 6H_2O \overset{\Delta}{\rightarrow} MgSO_3 + 6H_2O \uparrow$ (11)

$MgSO_4 \cdot 7H_2O \overset{\Delta}{\rightarrow} MgSO_4 + 7H_2O \uparrow$ (12)

The dry magnesium sulfite and magnesium sulfate crystals are calcined at 800 to 1000°C (1472-1832°F) in the presence of coke or a reducing atmosphere to regenerate the MgO and release SO_2. The calciner reactions are as follows:[3,4]

$MgSO_3 \overset{\Delta}{\rightarrow} MgO + SO_2 \uparrow$ (13)

$2C + O_2 \rightarrow 2CO \uparrow$ (14)

$MgSO_4 + CO \rightarrow MgO + CO_2 \uparrow + SO_2 \uparrow$ (15)

Regenerated magnesium oxide is moved to storage for reuse in the scrubber slurry system.

The SO_2 from the calciner is usually used as the feed stock for a sulfuric acid unit. However, depending on demands for sulfuric acid, it could be sold as liquified SO_2, liquified SO_3, or elemental sulfur; or used to enrich concentrated (98%) sulfuric acid to various grades of oleum.

Table 3-31 represents a material balance for this system, as applied to a 500-MW coal-fired plant.

3.4.2 Magnesium Oxide Scrubber Units - U.S.

Three full-size units have been operated in the United States. All used fuel with 2 to 2.5 percent sulfur content. They are as follows:

Utility name: Boston Edison
Unit name: Mystic Station, Unit No. 6
Unit location: Everett, Massachusetts
Unit size: 155 MW
Fuel properties: No. 6 fuel oil, 2.5% sulfur
Start-up date: April 1972
Note: Terminated June 1974

Table 3-31. MgO FGD SYSTEM MATERIAL BALANCE

Stream No.	1	2	3	4	5
Description	Coal to the boiler	Combustion air to air heater	Combustion air to boiler	Gas to the economizer	Gas to air heater
Rate, lbs/hr	375M	4,518M	4,075M	4,404M	4,404M
scfm		984M	888M	943M	943M
gpm					
Particulates, lbs/hr				39.3M	39.3M
Temperature, °F		110°F	535°F	890°F	705°F
Specific gravity					
Stream No.	**6**	**7**	**8**	**9**	**10**
Description	Gas to the ESP	Gas to the SO_2 scrubber	Gas to the reheater	Gas to the stack	SO_2 absorber slurry to tank
Rate, lbs/hr	4,848M	4,848M	5,156M	5,156M	14,217M
scfm	1,040M	1,040M	1,144M	1,144M	
gpm					26,085
Particulates, lbs/hr	39.3M	393			
Temperature, °F	310	310	127	174	127
Specific gravity					
Stream No.	**11**	**12**	**13**	**14**	**15**
Description	Makeup water	Total MgO feed	Liquor to slaker	MgO slurry to SO_2 abscrber	Slurry to thickener
Rate, lbs/hr	28.4M	16.9M	129M	146M	1,088M
scfm					
gpm	56.8		246	263	1,996
Particulates, lbs/hr					
Temperature, °F					
Specific gravity				1.11	

Table 3-31 (continued). MgO FGD SYSTEM MATERIAL BALANCE

Stream No.	16	17	18	19	20
Description	Thickener overflow	Recycle slurry	Thickener underflow	Liquor to centrate tank	Wet cake from centrifuge
Rate, lbs/hr	884M	14,013M	240M	137M	67M
scfm					
gpm	1666	25,745	340	260	
Particulates, lbs/hr					
Temperature, °F					
Specific gravity	1.06		1.2	1.06	
Stream No.	21	22	23	24	25
Description	Makeup water	Recycle to absorber	Feed to dryer	Dryer gas to cyclone	Cyclone dust to dryer
Rate, lbs/hr	23.4M	432M	73M	136M	5.5M
scfm				35M	
gpm	46.8	802			
Particulates, lbs/hr				5960	
Temperature, °F				400	
Specific gravity					
Stream No.	26	27	28	29	30
Description	Dryer gas to absorber	Dryer gas recycle to dryer	Dryer product to storage	0.1 to fire dryer	Air to fire dryer
Rate, lbs/hr	90.1M	45.8M	33.4M	3M	60.5M
scfm	23.2M	11.8M			13.2M
gpm					
Particulates, lbs/hr	159	318			
Temperature, °F			400		
Specific gravity					

Table 3-31 (continued). MgO FGD SYSTEM MATERIAL BALANCE

Stream No.	31	32	33	34	35
Description	$MgSO_3$ feed to calciner	Coke feed to calciner	Feed to calciner	Oil to calciner	Air to calciner
Rate, lbs/hr	210	40.1M	3.1M	3.1M	44.7M
scfm					9.7M
gpm					
Particulates, lbs/hr					
Temperature, °F					
Specific gravity					
Stream No.	**36**	**37**	**38**	**39**	**40**
Description	Calciner gas to cyclone	Calciner gas to bag filter	Cyclone dust to calciner	Bag filter dust to calciner	SO_2 to acid plant
Rate, lbs/hr	69.1M	61.1M	1719	187	134M
scfm	12.6M	12.6M			26.9M
gpm					
Particulates, lbs/hr	1910	191			4
Temperature, °F	1600				
Specific gravity					

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Table 3-31 Material Balance Notes

1. Calculations based on the following:
   - a. 105 percent stoichiometric magnesia
   - b. 9,000 Btu/kWh for the unit
   - c. 12,000 Btu/lb of coal
   - d. 3.5 percent sulfur coal (dry basis)
   - e. 92 percent of the sulfur in the coal evolves as SO_2
   - f. 14 percent ash in the coal (as fired basis)
   - g. 75 percent of the ash in the coal evolves as fly ash
   - h. 99.0 percent removal of particulates by the ESP
   - i. 90 percent SO_2 removal in the absorber
2. The letter M means 1,000.

Utility name: Potomac Electric and Power
Unit name: Dickerson No. 3
Unit location: Dickerson, Maryland
Unit size: 95 MW
Fuel properties: Coal, 2.0% sulfur
Start-up date: September 1973
Note: Terminated August 1975

Utility name: Philadelphia Electric
Unit name: Eddystone No. 1A
Unit location: Eddystone, Pennsylvania
Unit size: 120 MW
Fuel properties: Coal, 2.5% sulfur
Start-up date: September, 1975
Suspended operation in January 1976 pending relocation of calciner after acid plant shut down. Restarted in June 1977.

As discussed in the following sections dealing with individual units, SO_2 removal efficiency at all three locations has been over 90 percent. Both PEPCo-Dickerson and PECo-Eddystone have demonstrated over 99 percent particulate removal efficiency.

In general, however, all three units have experienced mechanical, material-related and product-related corrosion and handling problems. These have limited on-stream operation to between 27 and 80 percent of the total operating time of the generating units. The on-stream time was determined after an initial "shakedown" operating period. Boston Edison's Mystic Station Unit operated 80 percent or more of the time for three of the last four months of the test period.

3.4.2.1 Boston Edison - Mystic Station, Unit No. 6

System description - The flue gases from the boiler did not need to pass through any particulate removal equipment because oil was burned. To remove SO_2, flue gas was contacted cocurrently with an aqueous magnesium oxide slurry in a single-stage, venturi scrubber. A side stream from the scrubber slurry recirculating loop was constantly fed to a continuous centrifuge, and the aqueous phase from the centrifuge pumped back to the scrubber. The "wet cake" from the centrifuge was conveyed to a countercurrent, direct-fired dryer. The dry product was stored for later shipment to an off-site sulfuric acid plant, at Essex Chemical, Rumford, R.I., where the magnesium oxide was regenerated in a rotary kiln, and the resultant SO_2-rich gas stream used for sulfuric acid production as shown in Figure 3-53.[5,6]

Distinguishing Features - This unit was characterized by:

- Firing with No. 6 fuel oil during FGD operation.
- Fly ash from oil combustion was not removed before it reached the SO_2 absorber.
- The portion of the absorbing MgO slurry removed from the recirculating loop was not thickened prior to being fed into the centrifuge.
- The dryer and the calciner were both direct fired rotary kilns.
- There was no flue gas reheat prior to discharge through the stack.

Operating History - The average of the test runs for particulate removal in the single-stage venturi SO_2 scrubber was 57 percent as shown in Table 3-32. Resulting particulate emissions were low and averaged 36 kg/J (0.084 lbs/10^6 Btu).[7]

The scrubber did an excellent job of SO_2 removal. Test data indicate an average removal efficiency of over 91.6 percent. This was achieved at gas flow rates ranging from 12,036 m^3/min (425,000 acfm), which was the design value, to

Table 3-32. PARTICULATE REMOVAL - TEST RESULTS

BOSTON EDISON - MYSTIC STATION NO. 6[9]

Test No.	Boiler load, MW	Inlet gas rate		Particulates, lb/hr		% removal	Particulates	
		m^3/s	(acfm)	In	Out		Kg/J	(lb/10^6 Btu)
1	146	211.0	(446,953)	380	116	69.5	31.0	(0.072)
2	144	230.0	(486,991)	232	115	50.4	36.2	(0.084)
3	151	310.7	(658,207)	399	150	62.4	47.8	(0.111)
4	148	237.6	(503,233)	151	82	45.7	29.3	(0.068)

Table 3-33. SO_2 REMOVAL - TEST RESULTS

BOSTON EDISON - MYSTIC STATION NO. 6[10]

Test No.	Boiler load, MW	Inlet gas rate,		SO_2 in ppm - vol.	SO_2 out ppm - vol.	% SO_2 removal	SO_2 out	
		m^3/s	(acfm)				Kg/J	(lb/10^6 Btu)
1	146	211.0	(446,953)	926.1	71.1	92.3	59.0	(0.125)
2	144	230.0	(486,991	1004.5	89.0	91.1	93.9	(0.199)
3	151	310.7	(658,207)	983.9	63.3	93.6	95.9	(0.201)
4	148	237.6	(503,233)	833.3	86.6	89.6	114.7	(0.243)

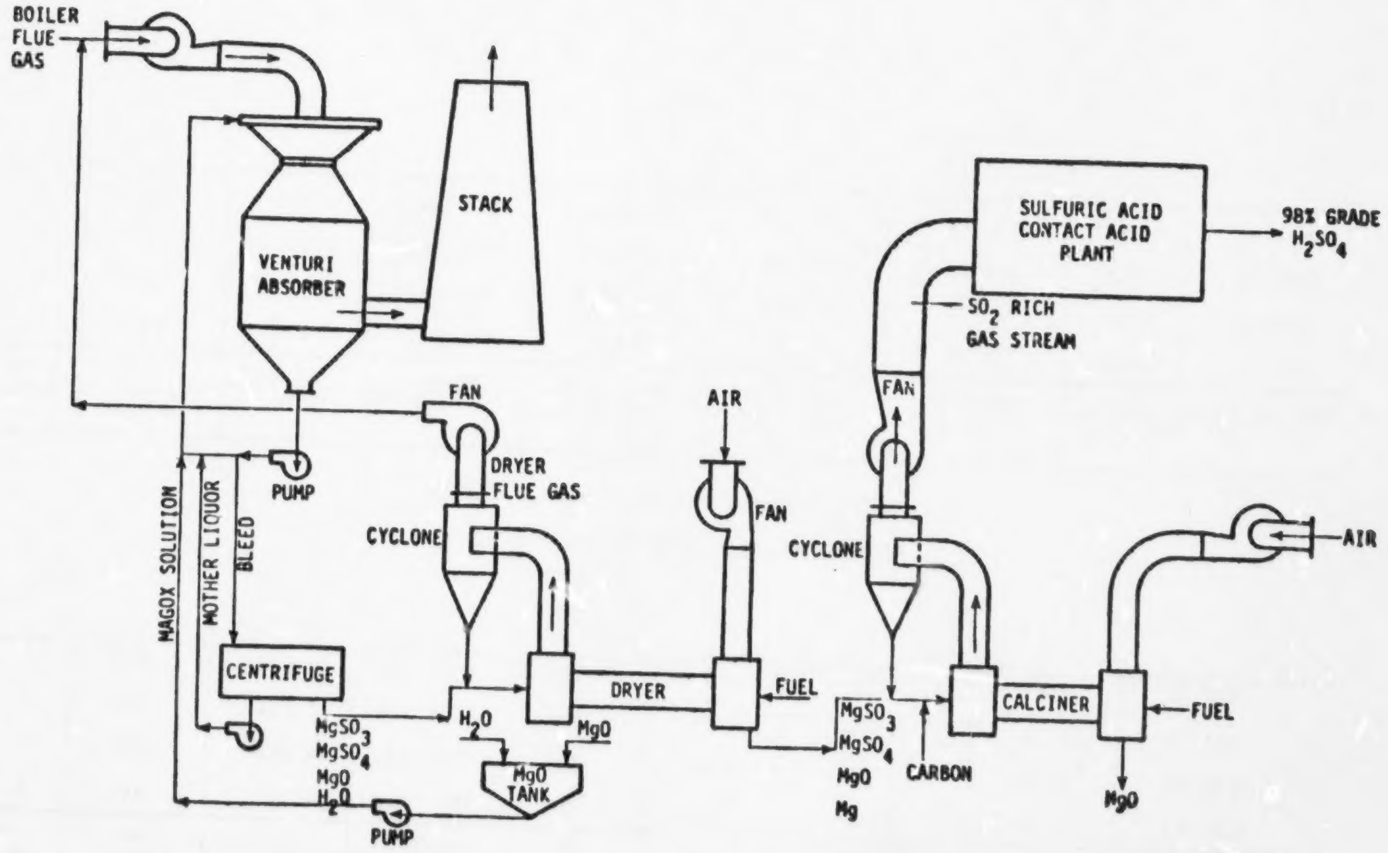

Figure 3-53. Mystic No. 6 FGD System and Essex Chemical Plant Regeneration Facility.

18,640 m^3/min (658,000 acfm), more than 54 percent in excess of the design value (Table 3-33). Outlet SO_2 concentrations averaged 82.7 kg/J (0.192 lb/10^6 Btu).[8]

The operability of the No. 6 Unit at Mystic for the entire test run is presented in Table 3-34. Operability is defined as hours of FGD operation divided by hours of boiler operation in a given period, expressed as a percentage. The unit worked best during its last four months, when operability was around 80 percent. It would have been approximately 85 percent but for a two-week outage of the off-site sulfuric acid plant. The FGD system had to shut down since MgO could not be regenerated.[11] From April 12 to May 10, 1974, the system achieved 100 percent operability.[12]

A correlation of SO_2 removal efficiency as a function of inlet SO_2 concentration and pressure drop across the scrubber based on actual test data is shown in Figures 3-54 and 3-55. The data show consistently high SO_2 removal efficiency for pressure drops over six inches of water. Although these data were obtained at the Mystic Station, the results should apply to other similar MgO systems.[13,14]

To maintain SO_2 outlet concentrations of less than 100 ppm, the pH of the recycle magnesium slurry stream leaving the absorber should be held between 6.8 and 7.5.[15]

If a regenerated magnesium oxide of maximum reactivity is required, the calcining kiln temperature and the control of the reducing atmosphere (through coke addition) are critical. The bulk density (specific gravity) of magnesium oxide is the best indicator of its activity. The desired bulk density is from 272 to 384 k/m^3 (17 to 24 lb/ft^3). Overburning the magnesium oxide produces unreactive periclase with a density as high as 480 to 721 kg/m^3 (30 to 45

Table 3-34. OPEPABILITY OF MgO SYSTEM AT MYSTIC NO. 6

Month/yr	Boiler operation (hr)	Scrubber operation (hr)	Operability %	Comments
Apr. 72 to May 73			17	The module operated intermittently because of mechanical difficulties. Operation of the magnesium sulfite crystals dryer presented a major problem.
Jun. 73	592	402	68	The longest period of continuous operation, 7.5 days, occurred during June and July.
Jul. 73	575	351	61	
Aug. 73		Boiler shutdown		The boiler was down for the annual overhaul. System operability decreased during the last quarter of the year because of heavy erosion/corrosion in the liquor recirculation pumps and centrifuge.
Sep. 73	637	243	38	
Oct. 73	627	377	60	
Nov. 73	629	162	26	
Dec. 73	658	86	13	
Jan. 74	555	152	28	Boiler-related problems caused frequent shutdowns in January and February.
Feb. 74	541	138	25	
Mar. 74	408	353	87	
Apr. 74	585	471	81	Two 7-day, continuous operation periods occurred during the month.
May 74	488	280	57	The 57-percent operability index value for the month was due to a 2-week outage of the acid plant rather than to FGD system failure.
Jun. 74	559	288	80	Demonstration program was completed and FGD system was shut down. EPA funding of $MgSO_3$ calcination has expired. There are no definite plans for restarting of this unit. Major problem areas encountered during the operation of this prototype unit included formation of trihydrate instead of hexahydrate sulfite crystal, dust problems in the dryer, lack of stack gas reheat, which caused condensation in the stack, louver damper problems, erosion of pumps, piping, and centrifuge, and minor ancillary equipment failures.
Test period completed				

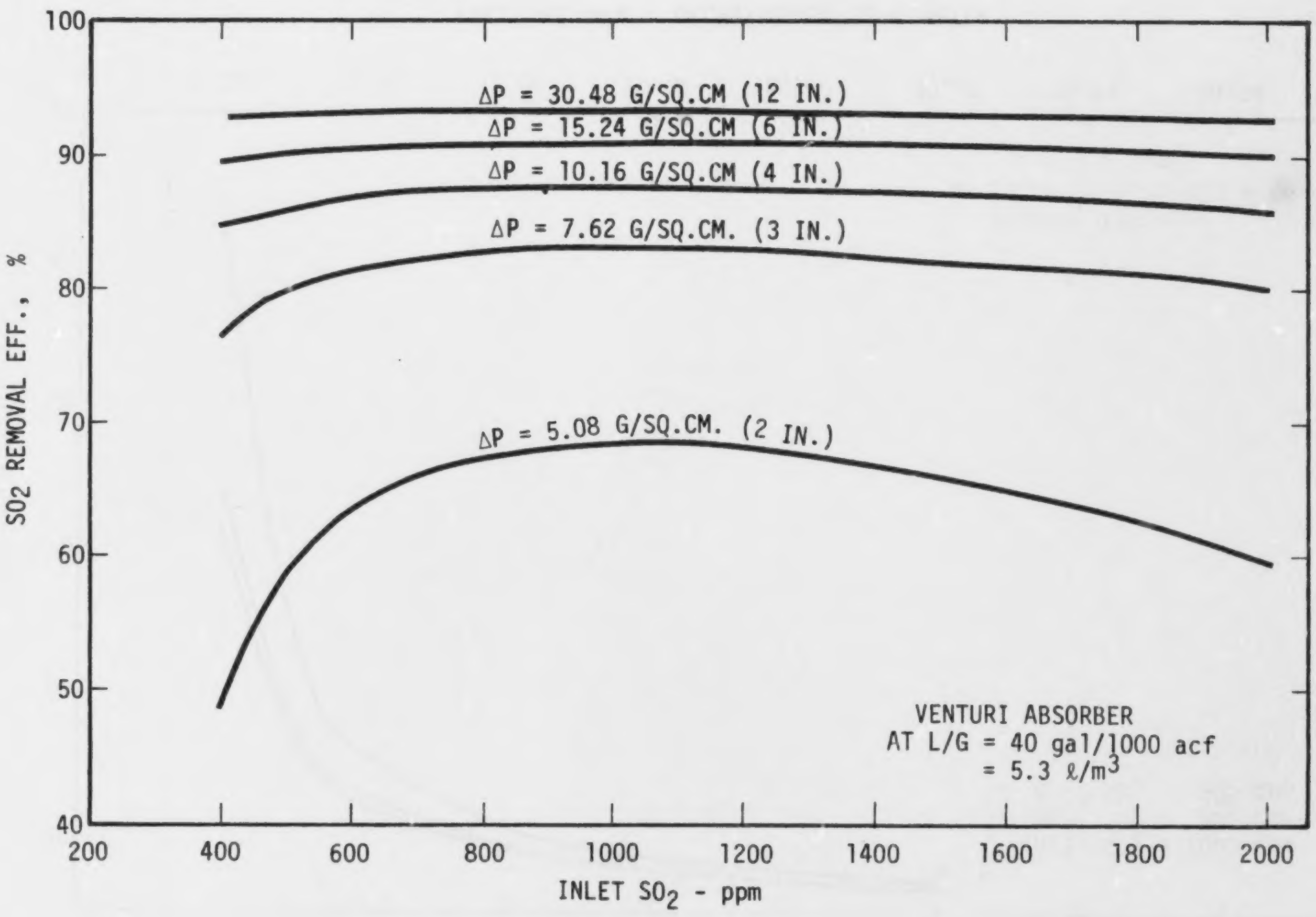

Figure 3-54. Effect of inlet SO_2 concentration and venturi pressure drop on SO_2 removal for the Mystic venturi absorber.[16]

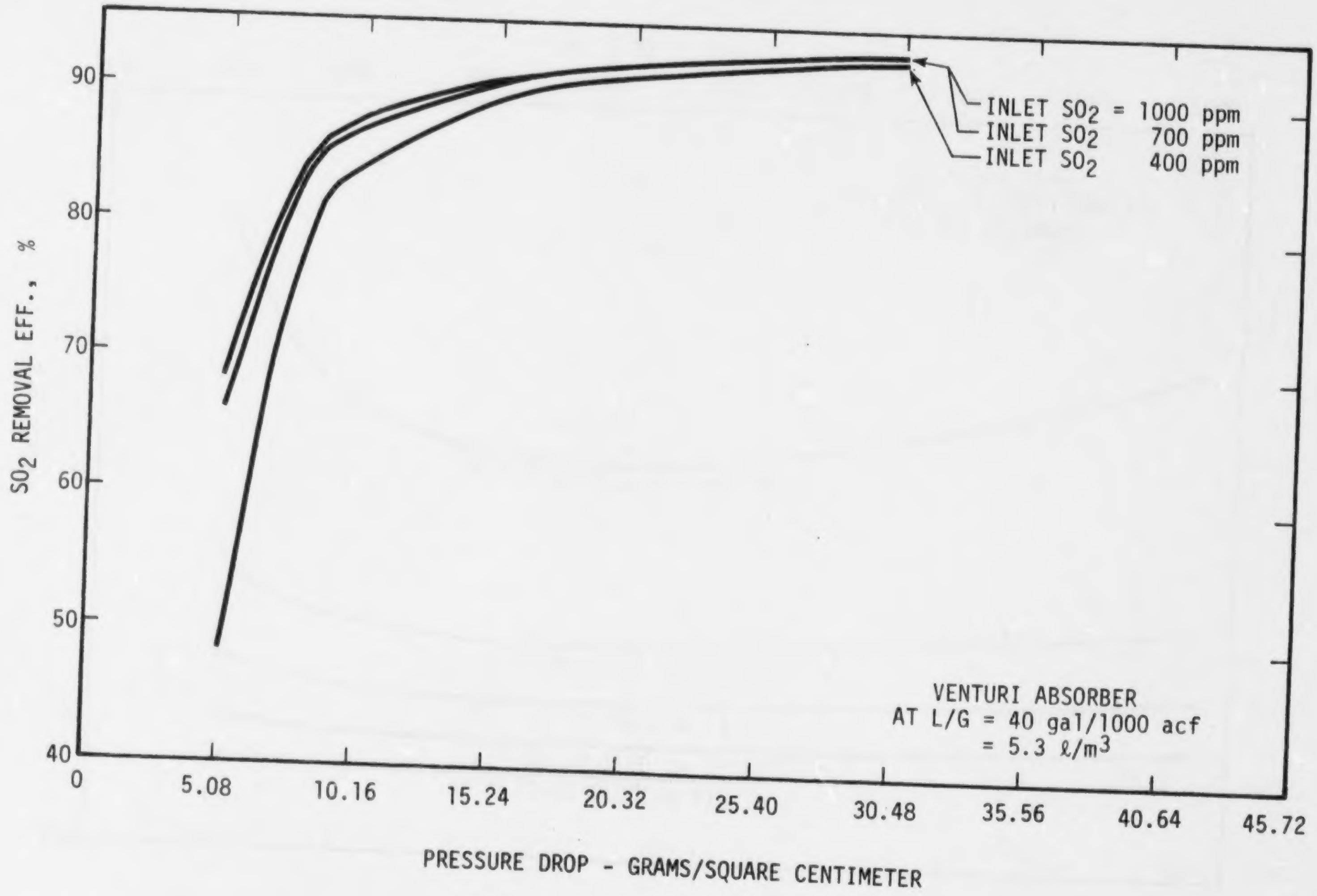

Figure 3-55. Effect of pressure drop on SO_2 removal for the Mystic venturi absorber.[11]

lb/ft^3).* Data from the test program show that formation of magnesia with low bulk density and high reactivity is promoted by lower temperatures (600-700°C; 1100-1300°F) and by increasing the carbon (coke) content (1.5-2%) in the calciner feed.[19]

Operating problems and solutions - The major problem at Mystic was the production of a high percentage of magnesium sulfite trihydrate ($MgSO_3 \cdot 3H_2O$), rather than the expected magnesium sulfite hexahydrate ($MgSO_3 \cdot 6H_2O$). The magnesium sulfite dryer was designed for operation with the relatively large hexahydrate crystal. Fine trihydrate crystals caused excessive dusting, solids buildup, loss of drying ability, and ultimately a number of major mechanical problems.[20] More than a year was required to accomplish the following:[21,22,23,24]

1) Modify dryer operation (trial and error)

2) Modify dryer internal configuration

3) Install hammers to prevent solids buildup on dryer walls

4) Change centrifuge wet cake feed-point

5) Install scalper and lump breaker to handle magnesium sulfite lumps leaving the dryer

6) Install pneumatic system for conveying material from the dryer off-gas cyclone dust collector to the storage silo. Dust was initially mixed with centrifuge wet cake; the mixture was "set up" in screw conveying equipment.

7) Vent dryer off-gas from cyclone dust collector back to the venturi scrubber inlet.

Erosion and corrosion of the carbon steel slurry recirculating pipes resulted in the adoption of rubber-lined pumps, valves, and piping throughout the SO_2 scrubbing system.[25,26,27,28,29]

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* A dead-burned, hardened form of magnesium oxide.

The centrifuge bowl plows, head plows, and conveyor surfaces underwent rapid wear, causing several shutdowns. Tests on stellite hard surfacing showed it could be used at least a year without repair. Wash connections were added to the centrifuge casing and an internal wash pipe was included to reduce or eliminate solids deposition.[30,31]

Handling regenerated magnesium oxide presented problems, including: 1) formation of overburned magnesia (periclase); 2) formation of oversized magnesia particles; and 3) difficulty in slaking the magnesia. Closer control of the coke (reducing atmosphere) and kiln temperature eliminated the periclase problem. Installation of a "scalper" for removal of oversized particles, and a pulverizer, reduced the number of oversized magnesia particles. Finally, raising the temperature of the magnesium oxide slurry tank to 82.2°C (180°F) increased the magnesia slaking rate.[32,33]

When low-sulfur oil was burned in the boiler during a test, the solids level in the recycle slurry stream from the centrifuge rose uncontrollably, forcing unit shutdown. This was caused by high levels of both magnesium sulfate and unreacted magnesia, resulting in increased viscosity and hindered settling. Use of low-sulfur fuel had changed the SO_2/O_2 ratio in the flue gas, generating greater oxidation and an increase in magnesium sulfate. Low-sulfur fuel was therefore abandoned.[34]

Summary - The operation showed that magnesium oxide scrubbing for sulfur oxide removal is technically feasible. The scrubber performed at or above the design efficiency

(90%) at gas flow rates 50 percent over design. Slurry solids separation was achieved and magnesium sulfate concentration controlled. No plugging and scaling occurred.[35]

Despite improvements, however, mechanical or equipment-related problems prevented the scrubber from meeting power plant needs, since it could not be operated for long, uninterrupted periods. Moreover, losses of magnesium oxide (20 to 50% makeup with virgin MgO, opposed to a design value of 5%) were unacceptable for much of the trial. In the last four months of operation, however, after numerous spills and leaks had been repaired and the calciner had been made to operate properly, losses were cut to 3.6 to 5 percent of the total magnesia feed.[36] This first-generation facility experienced significant mechanical and design problems, but many were solved by the end of the test program.

3.4.2.2 Potomac Electric and Power - Dickerson Station, Unit No. 3

System Description - This boiler burned 2-percent sulfur coal. Flue gas from the boiler normally passed through an ESP for fly ash removal before entering a two-stage venturi scrubber/absorber (Figure 3-56). The MgO system was tested both with and without the ESP, which can be bypassed.[37,38]

Particulates were removed in the first stage of a variable-throat venturi, operating with water. The gas was saturated and cooled to 49°C (120°F) in this stage.

Sulfur dioxide absorption was performed in the second stage by magnesium oxide scrubbing; slurry contacted the gas in a fixed-throat venturi.

The scrubber was designed to treat 139 m^3/s (295,000 acfm), which is half the output of a 190-MW boiler. After

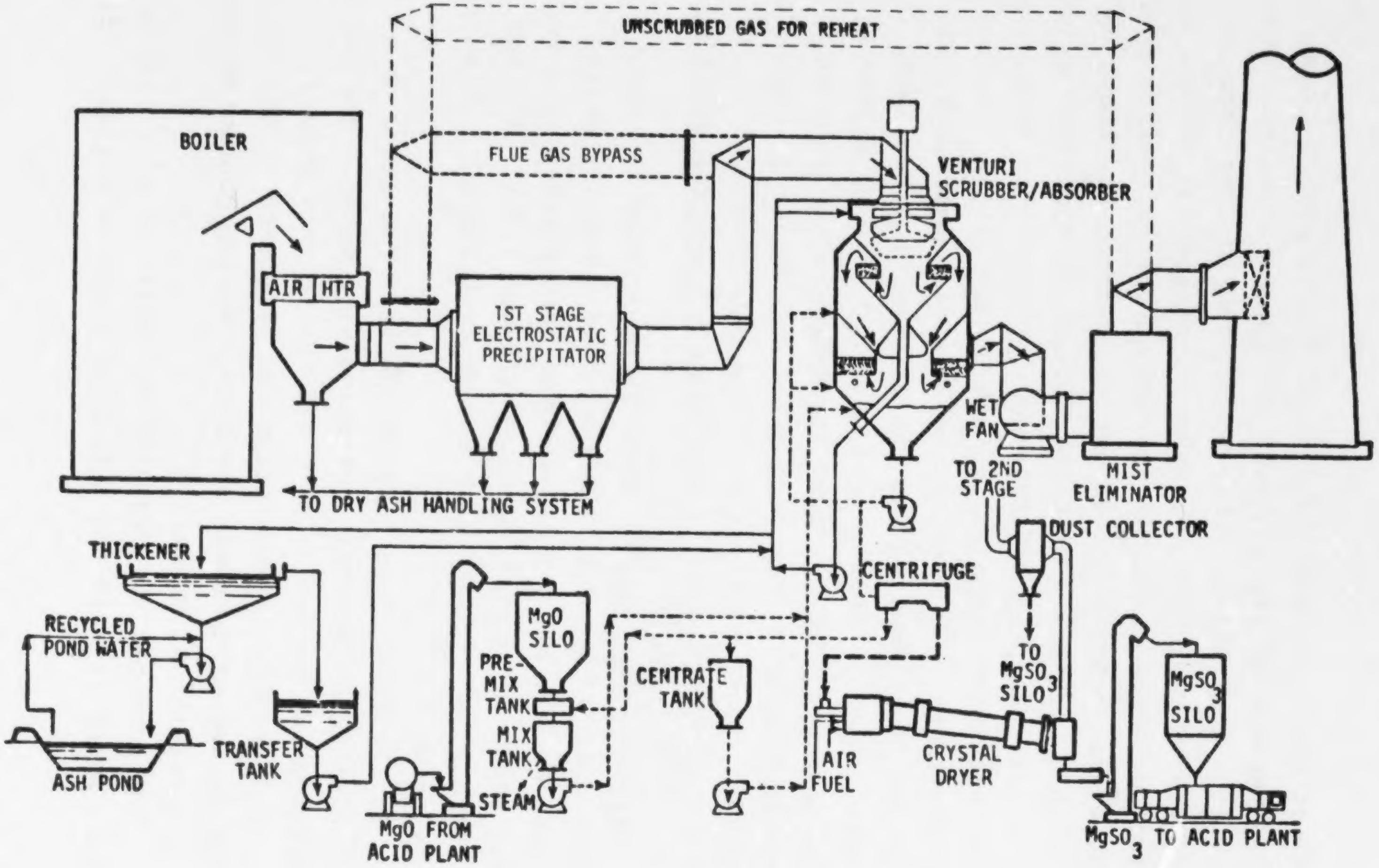

Figure 3-56. Dickerson No. 3 regenerative FGD system: General process diagram.

scrubbing, flue gas entered an induced-draft (wet) fan. It then went to a mist eliminator, where it was mixed with unscrubbed flue gas to raise the exit gas temperature to 70°C (170°F) before being vented through the stack.[39,40]

The liquid streams from the particulate and SO_2 scrubbers were separate. The particulate scrubber stage had two Alloy-20 liquid recycle pumps, each capable of moving the total flow. A side stream was continuously diverted to two thickeners of forty-foot diameter. Thickener overflow returned to the scrubber, together with any makeup water that might have been required. Thickener underflow was diluted and pumped to an ash pond.[41]

The SO_2 scrubber stream passed through a combination of three recycle pumps, each capable of pumping half the total slurry flow. A side stream was diverted continuously to a centrifuge.[42] The centrate (liquid phase) was pumped either to the magnesium oxide slaking operation or directly to the SO_2 scrubber section for recycling. From the centrigure, wet magnesium sulfite cake was screw conveyed to a cocurrent, direct-fired rotary dryer.[43,44] Dried cake went by bucket elevator to storage silos.

Dryer off-gases passed through a cyclone dust separator on route to the inlet of the second stage (SO_2 removal) of the scrubber. The magnesium sulfite collected in the cyclone went to the storage silos.[45,46]

The dry magnesium sulfite was transported from the storage silos to an off-site calcining/regenerating plant. Magnesium oxide was regenerated in a rotary kiln calciner and the SO_2 gas stream sent to a sulfuric acid plant.[47] The MgO regeneration unit and sulfuric acid plant at Rumford, R.I., were used for both the Mystic and Dickerson trials.

Unique Features of the Dickerson FGD Unit

- The unit was the first coal-fired utility application of the MgO FGD process.
- The particulate removal venturi and the SO_2 absorber venturi were both in the same shell.
- The magnesium sulfite dryer was an oil-fired rotary kiln using cocurrent flow.

Operating History - Dickerson Unit No. 3 achieved an average SO_2 removal efficiency of 88.9 percent during performance testing (Table 3-35). If the one low result were omitted, the average would be 90.4 percent.[48]

Concentrations of SO_2 ranged from 779 to 1419 ppm at the scrubber inlet. The unit was designed for 90-percent efficiency when burning 3-percent sulfur coal (equivalent to about 1850 ppm inlet SO_2).[49,50] At SO_2 inlet concentrations over 1000 ppm and at the design gas flow rate, removal efficiency was consistently above 90 percent. The L/G ratio was 5.350 l/m^3 (40 gal/1000 acf), and the pressure drop was 20 to 22 cm (8 to 9 in.) of H_2O. At lower pressure drop and lower inlet SO_2 concentration, reduced efficiency (82.9 to 88.7% removal) was noted. As pressure drop and/or inlet SO_2 concentration increased, removal efficiency increased.[51] Initially, the SO_2 removal was below design values (70% vs. 90%). The pressure drop across the absorber throat was then increased from 7.5 cm (3 in.) H_2O to 2.0 to 2.25 cm (8 to 9 in.) H_2O; thus SO_2 removal efficiencies in excess of 90 percent were achieved by better liquid/gas contact.[52,53] With regenerated MgO, SO_2 removal efficiency averaged 92 percent during tests in September 1973. This was at rated gas flow and SO_2 inlet concentrations from 771 ppm to 1419 ppm.[54]

Table 3-35. SO_X EMISSIONS TEST RESULTS

MgO FGD SYSTEM - DICKERSON[a]

Test series	ΔP,[b] cm (in.) H_2O	SO_2, ppm		SO_2 removal, %	SO_3, ppm	
		Inlet	Outlet		Inlet	Outlet
5A	38.4 (15.1)	779	78	90	34.6	3.56
5B	16.8 (6.6)	1373	157	88.7		
6	13.0 (5.1)	800	137	82.9	47.5	3.31
7	37.3 (14.7)	1418	88	93.9	2.9	0.64
8	13.2 (5.2)	1419	156	89.0	1.8	0.41

[a] Test results abstracted from York Research Corporation, Final Report, Y-8513, January 31, 1975.

[b] Second Stage (absorber) pressure drop.

Table 3-36. PARTICULATE EMISSION TESTS RESULTS FOR MgO SYSTEM AT DICKERSON

Test[a] series	Boiler load MW	ESP[b]	ΔP^c cm (in.) H_2O	Flue gas flow, 10^3 m^3/s (acfm)		Temp.		CO_2 %	O_2 %	Grams/nm^3 (Gr/scfd) 12% CO_2	Grams/am^3 (Gr/acf)
						°C	(°F)				
5A	183	Yes	25.9 (10.2)	141 (299)	Inlet	117	(242)	12.2	6.5	0.104 (0.190)	0.071 (0.131)
					Outlet	52	(126)			0.002 (0.002)	0.001 (0.002)
					Efficiency					99.0	98.5
5B	96	Yes	28.2 (11.1)	76 (161)	Inlet	116	(241)	11.9	6.7	0.055 (0.101)	0.038 (0.070)
					Outlet	47	(116)			0.003 (0.005)	0.002 (0.004)
					Efficiency					95.0	94.3
6	185	Yes	26.9 (10.6)	72 (153)	Inlet	116	(240)	12.3	6.4	0.065 (0.120)	0.046 (0.085)
					Outlet	58	(136)			0.004 (0.007)	0.004 (0.007)
					Efficiency					94.2	91.8
7	183	NO	27.2 (10.7)	133 (282)	Inlet	118	(245)	11.9	6.5	1.991 (3.654)	1.348 (2.474)
					Outlet	48	(118)			0.008 (0.014)	0.006 (0.011)
					Efficiency					99.6	99.6
8	176	No	25.4 (10.0)	60 (127)	Inlet	111	(232)	11.6	7.1	1.785 (3.276)	1.205 (2.212)
					Outlet					0.009 (0.017)	0.008 (0.014)
					Efficiency					99.5	99.4

[a] Average of four tests per series.

[b] Indicates whether the flue gas treated by the scrubber stage passed through the ESP.

[c] First stage (scrubber) pressure drop.

Table 3-37. REMOVAL EFFICIENCY FOR SELECTED PARTICLE SIZE RANGES

MgO SYSTEM - DICKERSON

Test Series	Removal efficiency, %		
	Above 5 μm	1 to 5 μm	Below 1 μm
5-1	99.0	97.5	94.6
5-2	99.8	99.2	99.0
5-3	99.0	99.9	98.6
5-4	98.8	97.5	97.0
Average	99.2	98.4	97.3

As shown in Table 3-36, particulate removal to below 0.02 grain per standard dry cubic foot (gr/scfd) could be attained, at any flue gas volume up to design flow rate, irrespective of whether the scrubber was accepting flue gas before or after it went through the ESP. Inlet loading to the scrubber ranged from 0.03 mg/m^3 (0.06 gr/scfd) to 1.99 mg/m^3 (3.66 gr/scfd). Scrubber efficiency was over 99 percent for particles over 5 μm; over 98 percent for particles between 1 and 5 μm; and over 97 percent for particles below 1 μm (Table 3-37).[55,56]

The operability of this MgO FGD system, defined as hours of operation divided by boiler operating hours, is shown in Table 3-38.

The unit was a prototype trial installation, built to obtain operating data. As such, equipment and materials were used that would not have been used in a long-term installation on a new plant. This operation fulfilled its purpose by showing areas where improvement was needed. The operability data in Table 3-38 show the downtime caused by mechanical and material failures.

Table 3-38. OPERABILITY DATA FOR DICKERSON NO. 3[57,58]

Time period	Operability, %
September 13, 1973 - January 14, 1974	27.4
December 9, 1973 - January 14, 1974	58.3
April 15, 1974 - May 1, 1974	40.5
August 1, 1974 - August 31, 1974	43.5
November 1, 1974 - November 30, 1974	44.6
November 15, 1974 - November 30, 1974	67.9
December 1, 1974 - December 31, 1974	57.9
Test Runs Completed August, 1975	

Problems and Solutions - The major equipment items (scrubber vessel, thickeners, centrifuge, and dryer) all performed well. Major problems were encountered, however, with construction materials for handling systems. Carbon steel pipe and slurry pumps were found to be inadequate against the corrosive/erosive properties of the slurry. Individual pump suction and discharge lines were necessary since leaks in pipes at or after a header could make installed spare pumps useless. The designs of the centrifuge discharge hopper, weigh belt feeders, and the dry magnesium sulfite bucket elevator were improved.[59]

Proper feeding of magnesium oxide to the slurry system became a problem, because of plugging in the mix tank and suction lines to the feed pumps. Later, there were difficulties with the proper slaking of regenerated MgO. The final solution to both problems was the installation of a steam-sparged, agitated premix tank/slaker to promote dissolution.[60,61,62]

Various rods, bellows and hangers, etc. corroded in the reaction vessel. It was found that the wrong materials had been used, particularly 304S.S. instead of the specified 316S.S. The proper materials were used to replace the corroded parts and this problem was solved.[63] Corrosion and erosion were severe in the recirculating piping of both first- and second-stage scrubber slurry. Fiberglass reinforced polyester (FRP) and epoxy were used to make repairs. For long-term commercial use, however, rubber-lined piping should be specified.[64]

Because the MgO regenerating facilities at the Essex Chemical sulfuric acid plant at Rumford, Rhode Island, were unavailable, the unit exhausted its supply of virgin magnesium oxide.* When the calciner became available, it was discovered that magnesium sulfite from Dickerson Station was predominantly hexahydrate, whereas that obtained from Boston Edison's Mystic Station was mostly trihydrate. This necessitated additional testing, to determine the proper operating parameters (temperature and feed rate) for calciner operation.[65,66]

The bucket elevator conveying the dried magnesium sulfite to the storage silo tended to overload and trip. The problem was traced to the discharge chute from the centrifuge, where wet magnesium sulfite cake tended to hang up and then break off in large chunks. Modifications to the discharge of the centrifuge outlet hopper and installation of larger buckets for the sulfite elevator helped overcome this.[67,68]

Problems affecting the centrifuge included buildup of material or wear inside the centrifuge, and changes in the physical/chemical form of the magnesium oxide. Use of washout ports and hardened steel surfacing, similar to those changes introduced at Mystic, should remedy this.[69]

The dryer was designed for use with chain sections, to allow better product drying and the reduction or elimination of product buildup.[70] The chains were never installed, however.

As is apparent from this summary, the major problems encountered at the Dickerson Plant were related to materials handling. They are now avoidable.

3.4.2.3 Philadelphia Electric-Eddystone Station, Unit No. 1A

System Description - This generating unit burns 2.5-percent sulfur coal.[71] The flue gas passes through a mechanical collector and an ESP before entering a system of three parallel scrubbing trains with a wet particulate scrubber, a reheater, and a scrubber booster (induced draft) fan in each train. A flue gas bypass system was also installed.[72] One of the three trains (Train C) has a SO_2 venturi scrubber in series with the wet particulate scrubber (Figure 3-57).

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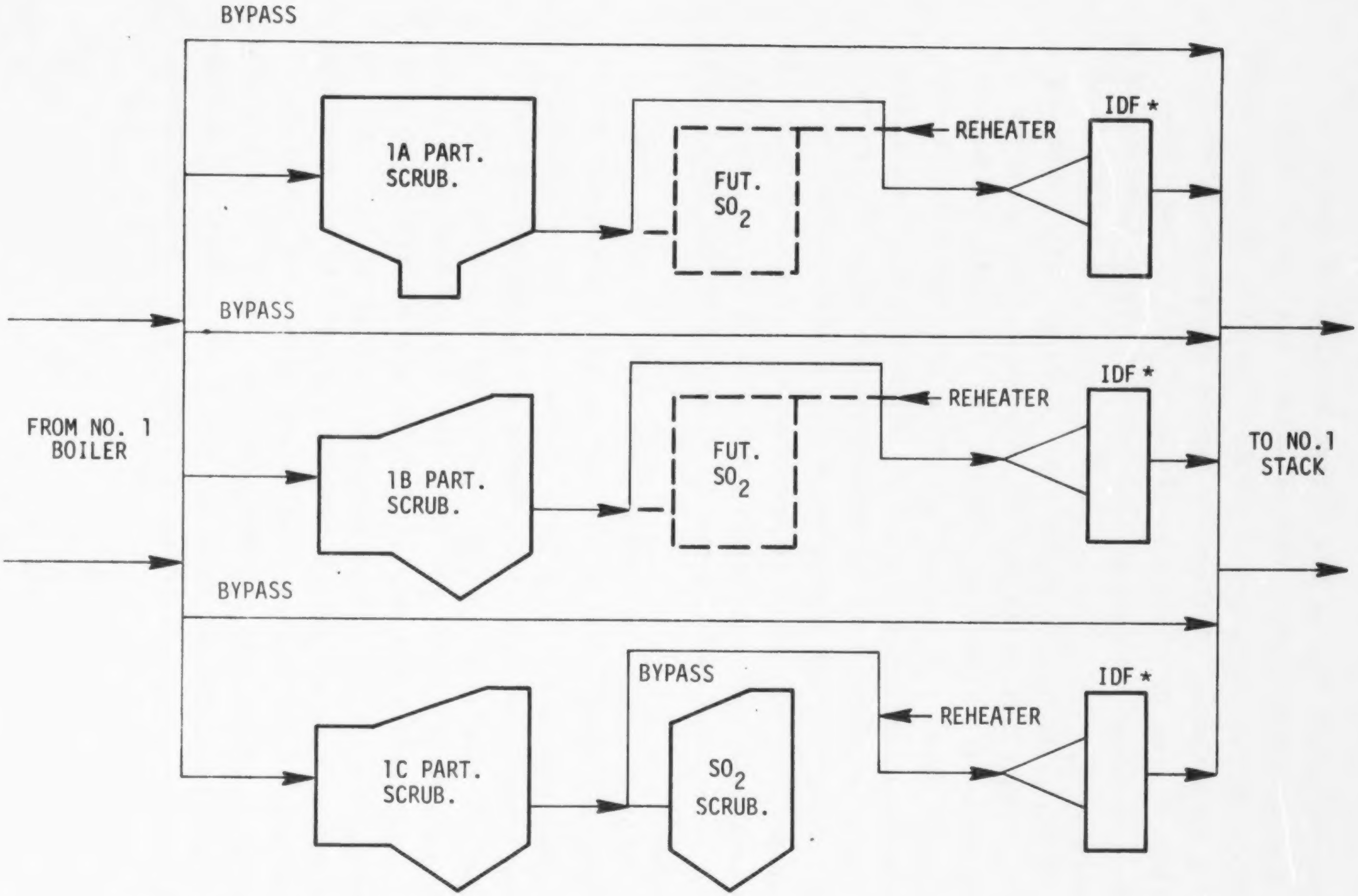
* The regeneration facility was being used to process $MgSO_3$ from the Mystic Plant at this time.

The particulate scrubbers are made of 316L stainless steel. The particulate and SO_2 scrubbers have separate surge tanks for slurry collection, pH control, and makeup. The surge tanks and the SO_2 scrubber are carbon steel with a polyurethane coating; the piping systems are rubber-lined.[73,74]

In the particulate scrubbing system shown in Figure 3-58, incoming flue gas first enters a venturi-type wet scrubber. The scrubber slurry goes to a surge/storage tank; the recirculating pump returns the slurry to the scrubber; and a continuous side stream is diverted to the waste water system.[75]

The gas stream is cooled from 146°C to 53°C (295-127°F), and humidified in the particulate scrubber. Humidified flue gas stream then enters the SO_2 venturi-type (Ventri Rod^R) scrubber (two stages of scrubbing and two stages of mist elimination), where the magnesium oxide/hydrate slurry scrubs out SO_2. The slurry then goes to the SO_2 scrubber surge tank, afterwhich the recirculating pump returns the slurry to the scrubber at a pH of 6.0. A continuous side stream is removed and sent to the magnesium sulfite recovery system; and makeup magnesium hydroxide slurry is fed to the SO_2 scrubber surge tank. The L/G ratio is 6689 liters per am^3/min (50 gal./1000 acfm); the scrubber operates at a pressure drop of 12 in. H_2O. Following SO_2



* INDUCED DRAFT FAN (IDF)

Figure 3-57. Eddystone unit 1A SO_2 removal system.

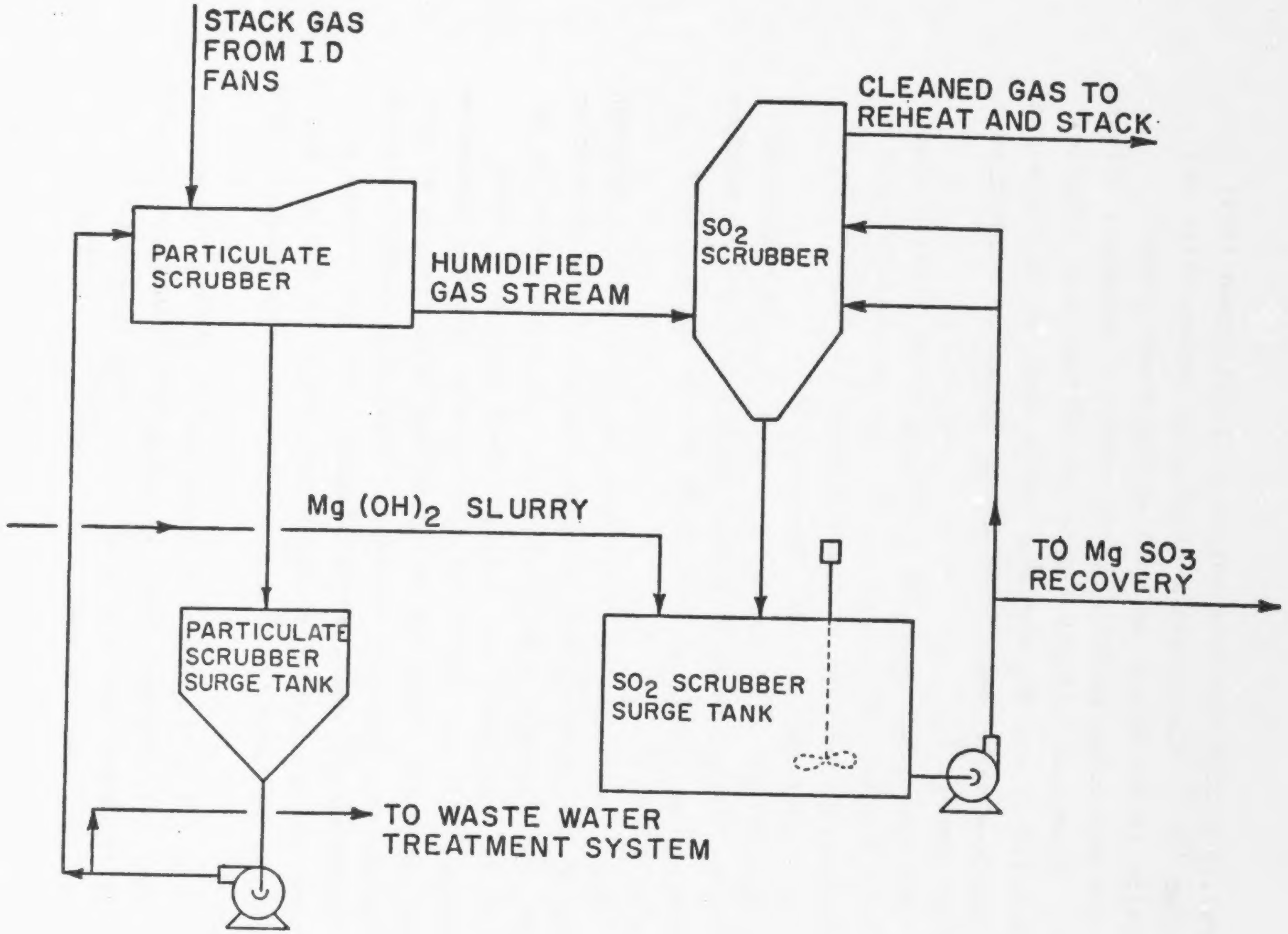

STACK GAS FROM I D FANS
PARTICULATE SCRUBBER
HUMIDIFIED GAS STREAM
SO_2 SCRUBBER
CLEANED GAS TO REHEAT AND STACK
$Mg(OH)_2$ SLURRY
PARTICULATE SCRUBBER SURGE TANK
SO_2 SCRUBBER SURGE TANK
TO Mg SO_3 RECOVERY
TO WASTE WATER TREATMENT SYSTEM

Figure 3-58. SO_2 scrubber system at Eddystone Plant.

removal, scrubbed gas then enters a direct-fired (oil) reheater.[76,77] The reheater eliminates condensation and corrosion in the stack, as well as the stack plume.

The magnesium sulfite slurry enters a thickener (Figure 3-59). Thickener underflow goes to a centrifuge; overflow is recycled to the SO_2 scrubber surge tank, or to the magnesium oxide slaking tank. Centrifuge liquor is handled in a similar manner.[78]

The centrifuge produces a wet magnesium sulfite cake, which is fed to a cocurrent dryer. The dry product is pneumatically conveyed to storage prior to shipment to an off-site regeneration plant (Figure 3-60).[79]

Exhaust gas from the dryer passes through a dust collector (collected dust is pneumatically conveyed to product storage), and on to the inlet side of the SO_2 scrubber (Figure 3-59).[80]

The off-site regeneration process uses a fluidized bed calciner for regeneration of magnesium oxide and liberation of SO_2 for sulfuric acid production. Magnesium sulfite is fed to the calciner from a solids feed tank by a screw conveyor. Hot air (954°C or 1750°F) and calcined magnesium oxide leave the fluid bed calciner and pass through an air preheater to a recovery cyclone. There the magnesium oxide is recovered and conveyed to storage. The gas stream is cleaned and scrubbed for removal of entrained MgO and any other particulates, before being piped to the sulfuric acid plant.[81]

Operating History - Although the Eddystone FGD system operated only a short time, it removed more than 90 percent of the SO_2 when both trays of ventri rods in the SO_2 absorber were used and when the L/G was 6700 ℓ/100 am^3/min

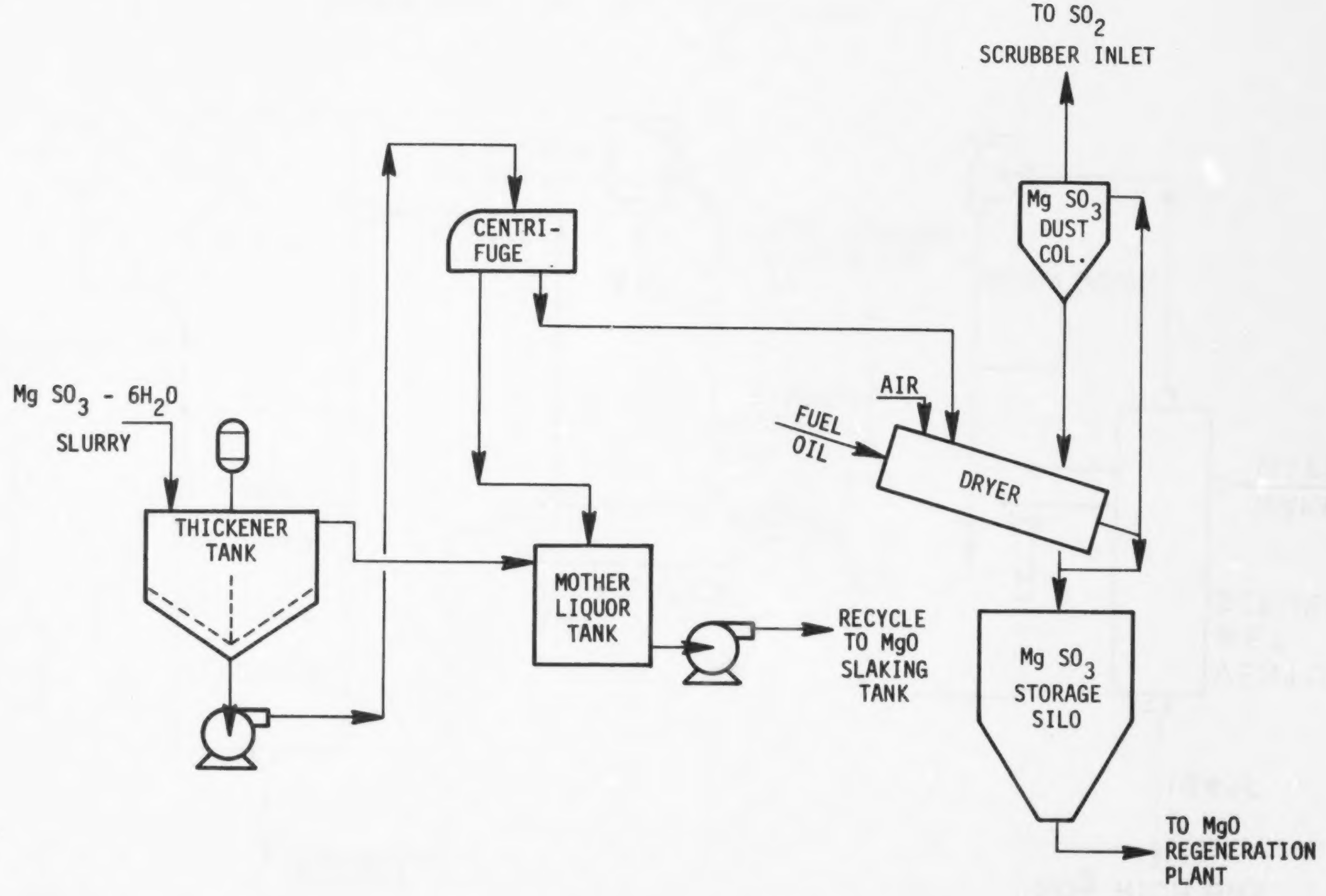


Figure 3-59. SO_2 scrubbing system - Mg SO_3 recovery.

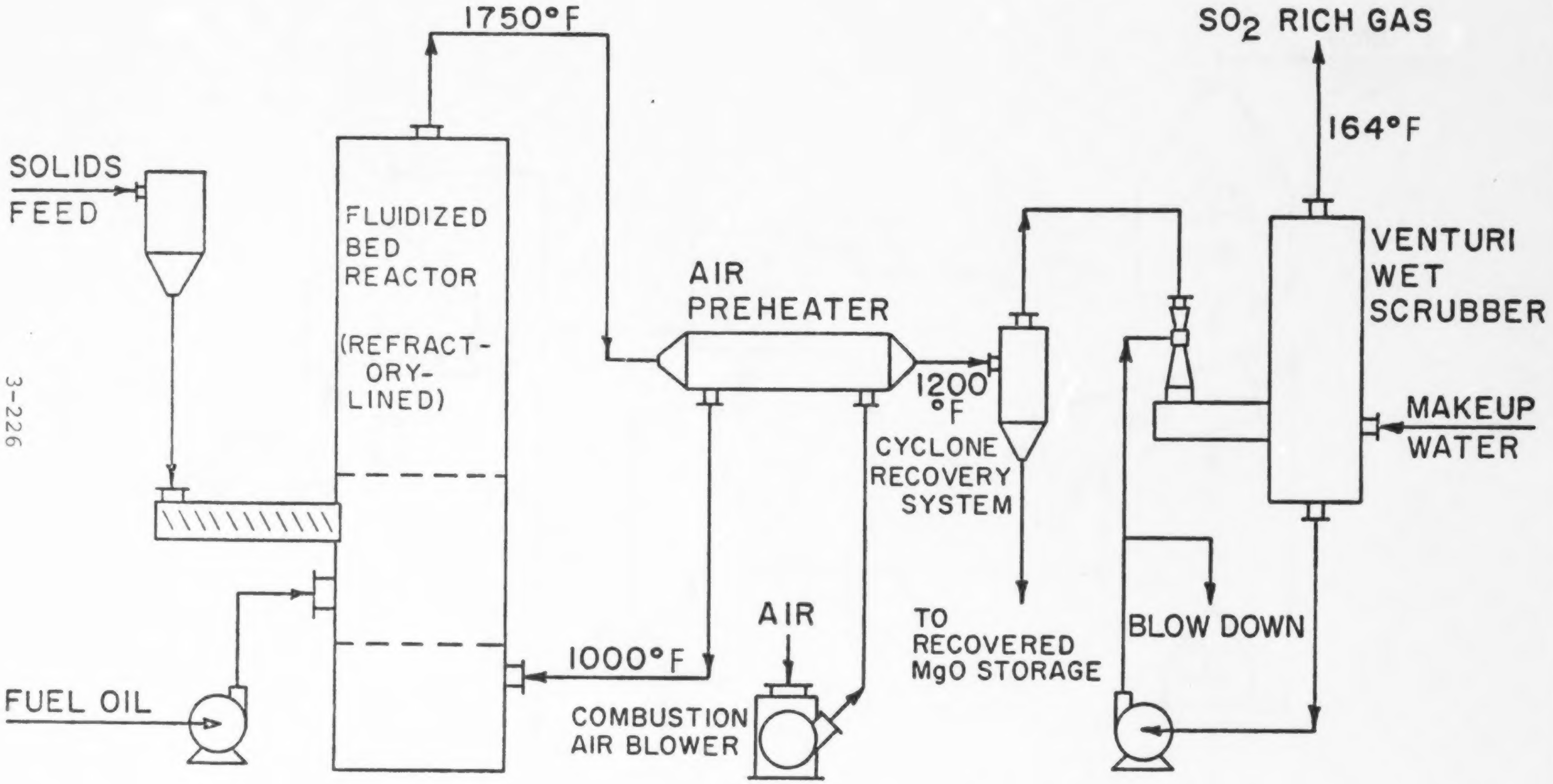


Figure 3-60. MgO regeneration plant.

(50 gal./100 acfm). Eddystone must meet an SO_2 emission standard of 0.26 g/MJ (0.6 lb $SO_2/10^6$ Btu). Published test data are not available.

Limited operability data given below show the initial start-up difficulties.[82] Sulfur dioxide absorption ceased from January 1976 until June 1977 while the regeneration facility was being moved.

Particulate Scrubbers Operability

1C: July 23, 1975 - January 31, 1976: 70 percent
1B: August 15, 1975 - January 31, 1976: 55 percent
1A: October 2, 1975 - January 31, 1976: 24 percent

SO_2 Scrubber

1C: October 2, 1975 - December 31, 1975: 33 percent

The SO_2 scrubber was shut down because Olin Chemical closed the acid plant serving the MgO calciner. The regeneration equipment was relocated to Essex Chemical, Newark, N.J., and restarted in June 1977.

Problems and Solutions - The particulate scrubbers experienced many problems. All three scrubber booster fans developed high shaft vibration (1B fan failed in November 1974, 1C failed in July 1976, and 1A failed in September 1976). These failures were caused by excessive clearance of the slip fit between the shaft and the wheel hub, resulting in high vibration at operating temperatures. This construction error meant that all three fans had to be returned to the manufacturer for shrink-fit repair. They were back on line in March 1977.[83,84]

The recirculation pumps for the particulate scrubber were too small. This problem, too, was resolved with the manufacturer.[85]

The particulate scrubber by-pass dampers are operated by a pneumatic cylinder through a lever arm. The lever arms failed several times, so heavier ones with longer stroke pneumatic cyclinders have been installed.[86]

The particulate scrubber suffered extensive corrosion caused by pH levels as low as 0.5, chlorides of over 200 ppm, and the "shock" of a "hot start" procedure. The problem has been lessened by a) the addition of caustic to maintain the system pH at 2.5; b) higher blowdown to control the chlorides; and c) the introduction of a "cold start" procedure.[87]

Within 3 months the polyurethane coating in the 1C SO_2 scrubber and particulate surge tanks began to blister and peel. All the vessels were sandblasted and coated with "flake glass" polyester coatings. This has held up well.[88]

The reheaters on all three systems developed refractory problems, caused by faulty flame/torch placement, the failure of ultraviolet scanners, and inadequate heat dissipation. These problems have all been resolved.[89]

In the most recent test period (May 1977 to August 1977) the main problems were with ancillary equipment (rotary valves, pumps, etc.). Vendors are working with United Engineers and PECo personnel to resolve these problems. The major units (scrubbers, absorber, etc.) worked properly.

3.4.3 Description of Japanese MgO Scrubber Installations

Japan has three industrial MgO FGD installations (Table 3-39), none of which operates solely on boiler flue gas. Brief descriptions of them are given in this section. Although these units do not operate on boiler flue gas, their operating principle is identical to that of systems on coal-fired power plants in the United States, and the equipment used is very similar. The Japanese plants reportedly have worked well; specific operability data have not been published, however.

Table 3-39. MgO FGD SYSTEMS IN JAPAN[90,91]

Process developer	User	Plant site	Gas flow 1000 nm^3/hr	Gas source	Inlet SO_2 ppm	Installed	By-product
Onahama-Tsukishima	Onahama Smelting	Onahama	84 (28 MW)*	Copper smelter	25,000	1972	H_2SO_4 240 T/D
Chemico-Mitsui	Idemitsu Kosan	Chiba	500 (162 MW)*	Claus & boiler	(Not available)	1974	S
Mitsui Mining (shut down)	Mitsui Mining	Hibi	80 (25 MW)*	H_2SO_4 plant	(Not available)	1971	H_2SO_4 18 T/D

* Equivalent MW based on gas flow from oil-fired boilers.

3.4.3.1 Onahama - The Onahama Smelting operation cleans the gas stream from two reverberatory copper smelting furnaces. Half the gas stream is cleaned using a MgO FGD system, the other half by a lime scrubber system.[92] The MgO scrubber is not currently operating, because of the reduced operation of the smelter and the increased demand for gypsum (the by-product of lime scrubbing). For several years, however, the lime and MgO systems operated in parallel (Figure 3-61)).

The furnace gas, containing 2.0 to 3.0 percent (20,000 to 30,000 ppm) SO_2, enters waste heat boilers, passes through an ESP and a water wash system, and is further cooled by surface coolers. At this point, the cooled gas stream is divided and routed to the MgO or the lime FGD system. The MgO system uses two turbulent contact SO_2 absorbers (TCA). The primary product is $MgSO_3 \cdot 6H_2O$. The two MgO absorbers operate alternately every two weeks. These TCA absorbers originally were backwashed with sulfuric acid for removal of magnesium sulfite scale; more recently, however, only water has been used. The problem of wear on the polyethylene balls has necessitated their replacement every few months. The absorber off-gas passes through a mist eliminator and an ESP, to be discharged from the stack at 60°C (140°F).[93,94]

The $MgSO_3$ stream from the scrubber enters a liquid cyclone; the resulting concentrated slurry is centrifuged and the $MgSO_3$ cake that is produced is moved by screw conveyor to a rotary steam dryer. Centrifuge filtrate is returned to the scrubber loop. The $MgSO_3$ crystals that are formed are normally about 200 μm (7.87×10^{-3} in.) in size. In the drying operation, less than 3.5 kg of steam/kg SO_2 is consumed. Both dryer and centrifuge operations have been

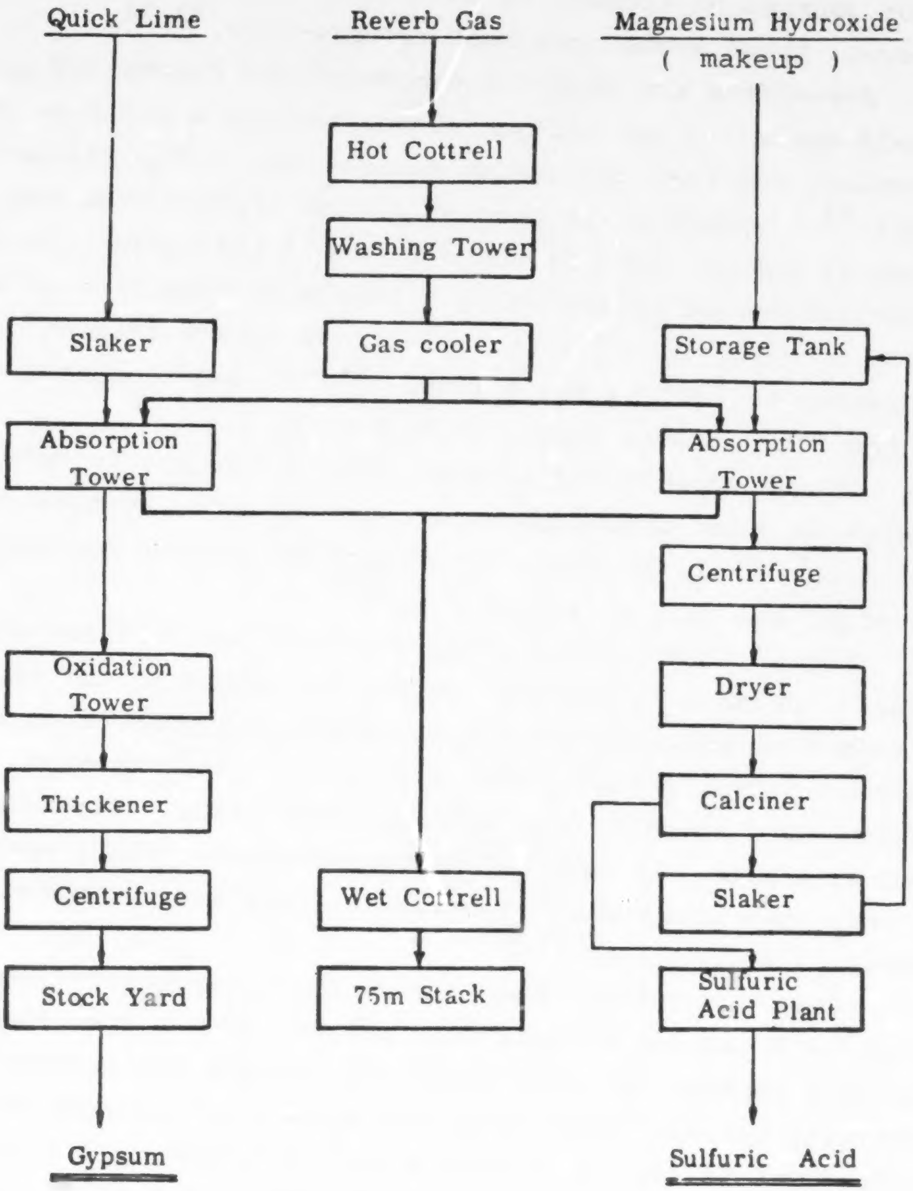


Figure 3-61. Flowchart - Onahama reverberatory furnace SO_2 control system.

major sources of trouble in the United States; in Japan, however, these areas have been trouble-free.[95,96]

Steam from the dryer is condensed and discharged as liquid waste. Dryer off-gas passes through a baghouse and scrubber, and then through an exhaust fan to the discharge stack.[97] Typically, 10 percent of the product from the dryer is $MgSO_4$. Coke is added to the dried product through a variable-speed rotary valve. This aids reduction of the $MgSO_4$ to $MgSO_3$ in a direct, oil-fired, rotary calciner operating at 750° to 800°C (1382-1472°F) and consuming oil at the rate of less than 0.13 kg/kg SO_2.

Of the regenerated product, only 1 percent is $MgSO_4$. Magnesium oxide agglomerates are ground prior to slaking. The material in the slaker comprises 92 percent regenerated material and only 8 percent virgin MgO.

Losses of MgO occur primarily from the calciner exhaust. On an average, the proportion of SO_2 in the exit gas stream from the calciner is 11 percent (10 percent to 13 percent normal range). Calciner gases, after washing, become feed stock to the sulfuric acid plant. Regenerated MgO is slaked and recycled to the absorber. Plant operational specifications and component data are presented in Tables 3-40 and 3-41.[98,99]

This FGD system demonstrated SO_2 removal efficiencies from 99.5 percent to over 99.8 percent from a concentrated (2 to 3 percent SO_2) reverberatory furnace gas stream [utility boilers fired with coal with a 3.5 percent sulfur content would typically have a concentration of 0.2 percent SO_2 (2000 ppm]. The design called for 99.5-percent removal from a 3 percent inlet gas stream. As a safety factor, the smelter must be operated 50 percent to 70 percent in excess

Table 3-40. MgO PLANT OPERATIONAL SPECIFICATIONS - ONAHAMA SMELTER

Reverberatory gas:	
Volume (dry)	1500 nm^3/min (52,950 cfm)
SO_2	2.4 to 2.6 percent
O_2	6.0 percent
CO_2	12.0 percent
N_2	Balance
H_2	0.12 (kg/kg dry gas)
Temperature	300°C, 572°F
Pressure drop (MgO Plant)	8.0 (in. H_2O (20.3 cm)
Calciner (rotary kiln) gas:	
SO_2	10 to 13 percent
Tail gas:	
Volume	270 nm^3/min (9531 cfm)
SO_2	<100 ppm (≈20 ppm)
Steam consumption at dryer	<3.5 kg/kg SO_2
Oil consumption at calciner	<0.13 kg/kg SO_2
$Mg(OH)_2$ makeup	<0.08 kg/kg SO_2

Table 3-41. MAJOR COMPONENTS OF MgO PLANT - ONAHAMA SMELTER

Absorption step:	
TCA tower	4.0 m (13.2 ft) dia. x 26.5 m (87.2 ft) length x 2 (1 backup)
Blower	2200 nm^3/min (77,660 cfm) x 900 Kw x 2 (1 backup)
Separation and drying step:	
Centrifuge	Essherwise type 10
Dryer	900 m^2 (9720 ft^2) x 2 [3 m (9.8 ft) diameter, 25m (82 ft) long]
Thermal decomposition step:	
Rotary kiln	3.4 m (11.2 ft) dia. x 52 m (171 ft) length

of the 99.5 percent goal. Operation, therefore, is strictly controlled to remove 99.7 percent of the sulfur input. The design outlet SO_2 emission was 100 ppm. Actual operation has achieved less than 50 ppm.[100,101,102,103,104]

3.4.3.2 Idemitsu Kosan - At Idemitsu Kosan's Chiba refinery, Mitsui Miike Machinery Co. constructed a MgO FGD system to clean 500,000 nm^3/hr (294,167 cfm) of waste gas from industrial boilers and Claus furnaces. The unit went on stream in November 1974 and has consistently achieved 95 to 97 percent SO_2 removal efficiency at an inlet SO_2 concentration of 2,500 ppm.[105,106]

The venturi particulate scrubber and venturi SO_2 absorber are both in the same shell. The $MgSO_3$ slurry discharge from the absorber is pH-adjusted and filtered. The $MgSO_3$ cake is dried in a rotary dryer countercurrent to the hot gas stream; the dried $MgSO_3$ is calcined in an oil-fired, rotary kiln. The $MgSO_3$ cake contains 10 to 15 percent $MgSO_4$; coke is added in the calcination step to reduce the sulfate to sulfite. Regenerated MgO is slaked and then wet-milled to increase its reactivity. Off-gas from the calciner goes through a cyclone, a wet venturi, and a wet precipitator. This cleaned gas, containing 10 to 12 percent SO_2, is fed to the Claus furnace, where H_2S recovered in the refinery is reacted with SO_2 to produce elemental sulfur.[107]

The form of magnesium sulfite, trihydrate, or hexahydrate is an important key to plant operation. Usually hexahydrate is preferred, because it grows in much larger crystals than does trihydrate. Mitsui Miike has found that mixing the two forms gives the best results. There has been no problem with dryer operation. The major problem, however, was with the the firebricks in the rotary kiln. They became dislodged, possibly the result of cooling by rain. To prevent corrosion, chloride concentration in the scrubber liquor is maintained below 2,300 ppm. Normally wastewater is retained because of the low inlet chloride concentration.[108,109]

3.4.3.3 Mitsui Mining - In 1971, Mitsu Mining and Smelting Co. built an MgO system at the Hibi Works. This unit treated 80,000 nm^3/hr (47,067 cfm) of tail gas with 1500 ppm to 2000 ppm SO_2 from a sulfuric acid plant. A cross-flow absorber was used, and $MgSO_3 \cdot 6H_2O$ was formed. Sulfite crystals were separated by centrifuge from the solution, which contained some magnesium sulfate. The sulfite was dried in a rotary dryer and then calcined in an indirect-fired calciner at 750°C. The concentrated SO_2 was returned to the acid plant. Magnesium sulfate solution was concentrated to produce $MgSO_4 \cdot 7H_2O$, which was sold for fertilizer and other uses.[110]

3.4.4 Engineering Design Parameters

Sulfur dioxide collection efficiencies in excess of 90 percent can be achieved with a MgO FGD system on coal fired boilers. Key design and operating parameters for 90 percent removal include:

- Operation of venturi scrubbers at pressure drops of 2.5kPa (10 in. H_2O) or greater, with an L/G ratio of 40 to 50; use of a TCA scrubber operating at 2.0 kPa (8 in. H_2O) on smelter reverberatory furnace off-gas.
- Maintenance of slurry pH (measured in the absorber discharge tank) in the range of 6.0 to 7.5 (Figure 3-62).
- Slaking (dissolution) of MgO in water in a heated, agitated tank before it is added as makeup to the absorber.

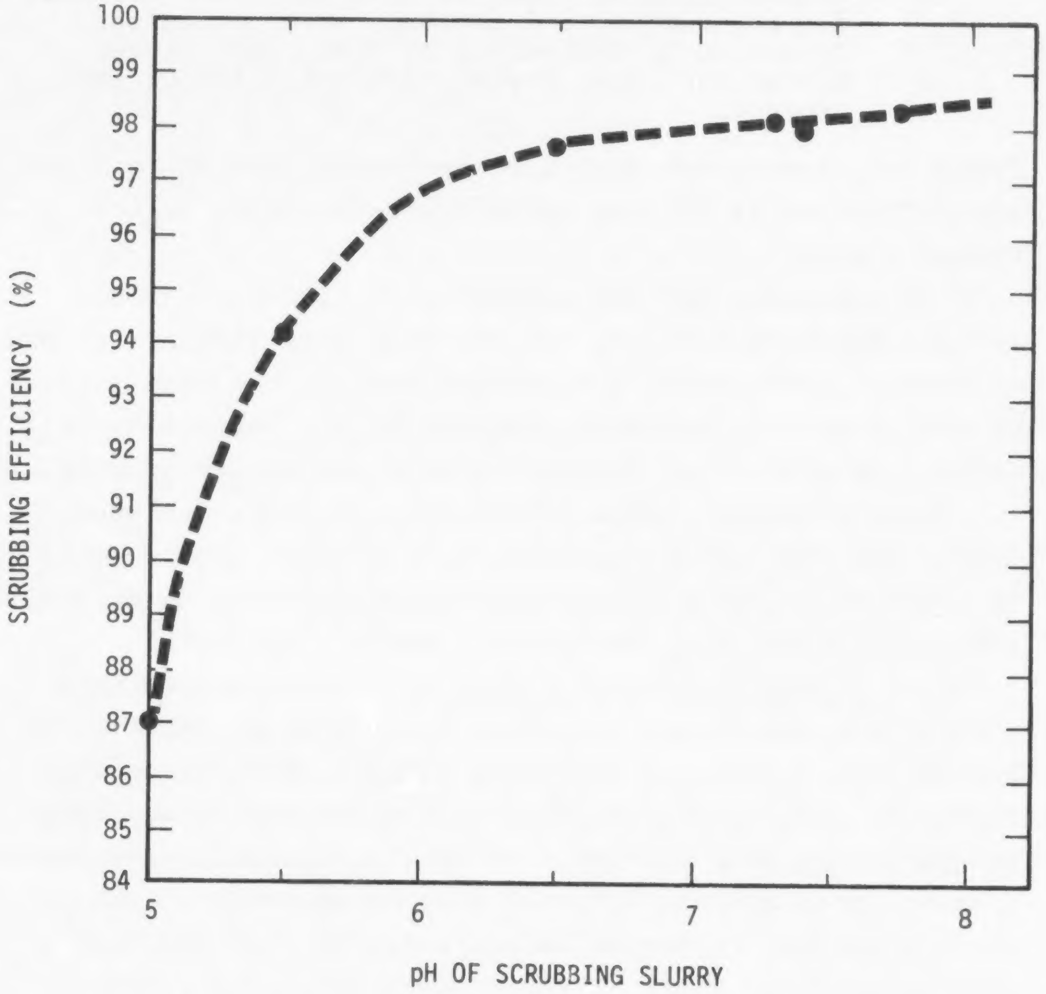


Figure 3-62. The affect of pH on SO_2 scrubbing efficiency.[111]

- Avoidance of over-calcining or dead-burning of regenerated MgO, to maintain high reactivity.
- Prescrubbing of flue gas with water for residual fly ash removal and cooling, which minimizes formation of trihydrate of $MgSO_3$ and insures better handling, drying, and calcining of the product.

Table 3-42 summarizes operating parameters and SO_2 collection efficiencies for the three plants operated in the United States.

In the past, MgO FGD systems installed by Chemico (Mystic and Dickerson) and United Engineers (Eddystone) had no overall performance guarantees; rather, the manufacturer of each piece of equipment guaranteed that particular unit against manufacturing defects, guarantees varied greatly.

Now, however, Chemico is willing to guarantee the entire MgO FGD system mechanically, as well as to specify that the unit will meet the applicable SO_2 regulation 0.52 g/MJ (1.2 lb/10^6 Btu, 90 percent removal, etc.).[112]

3.4.4.1 Applicable Research Studies - Chemico sponsored work at the New Jersey Institute of Technology (NJIT), Newark, New Jersey, to determine conditions favoring the formation of magnesium sulfite trihydrate and hexahydrate. Trihydrate is more difficult to handle physically and has a smaller, more granular crystal than hexahydrate. (The primary product at Boston Edison's Mystic unit was the trihydrate form; while at PEPCO Dickerson, hexahydrate predominated.)

The NJIT study of SO_2-bearing gas (at 160°C, 320°F), with MgO slurry at 49°C (120°F), 57°C (135°F), 65°C (149°F) and 80°C (176°F), showed that at 49°C (120°F) and 57°C (135°F) the major product was $MgSO_3 \cdot 6H_2O$, while at 65°C (149°F) and 80°C (176°F) it was the trihydrate $MgSO_3 \cdot 3H_2O$.

Table 3-42. SUMMARY OF MgO SCRUBBING SYSTEM OPERATING PARAMETERS

Plant	ppm	Inlet SO_2 lb/10^6 Btu	g/MJ	Outlet SO_2 ppm	SO_2 removal efficiency, percentage	ΔP		L/G		Slurry pH
						KPa	in. H_2O	ℓ/1000 m^3	gal/1000 acfm	
Mystic (Boston Edison)	1042	1.05	0.452	89	91.5	2.0	8	4400	33	7.1
Dickerson (Potomac Electric)	1092	1.60	0.689	121	88.9	2.0	8	5350	40	7.1
Eddystone (Philadelphia Electric)	1500	1.90	0.819	Below 150	>90	2.5	10	6700	50	6.3

The presence of magnesium sulfate in solution (up to 18%), and the passage of excess air, had no significant influence on the nature of the hydrate formed.[113]

Chemico also conducted research at the MgO regeneration facility in Rumford, Rhode Island, where the reactions of SO_2 at 149°C (300°F) with both virgin and regenerated MgO were studied at 57°C (135°F), 60 to 63°C (140 to 145°F) and 66°C (150°F). The pH was maintained at 6.8 to 7.2. Samples were collected periodically and the filtrates were returned to the reactor. The results of the study agreed with those performed at the NJIT; i.e., hexahydrate predominated at the lower temperatures.[114]

At 57°C (135°F) the major product was the hexahydrate; trihydrate was the predominant product at 66°C (150°F). The study at 60 to 63°C (140 to 145°F) was interesting--it showed the gradual change from hexahydrate (original product) to trihydrate (final product). The complete transformation from hexahydrate to trihydrate required 4 to 5 hours. The study at 60 to 63°C (140 to 145°F) showed also that the transition of hexahydrate to trihydrate was accompanied by a reduction of crystal size (from 200 to 50 μm).[115]

Analysis of centrifuge cakes from PEPCO and Boston Edison supported the contention that: a) the ratio of the two hydrates at any temperature is controlled by the equilibrium value at the temperature; b) hexahydrate persists even above the transition temperature; and c) the rate of transition is slow.[116]

Further work was done to determine the drying mechanism of magnesium sulfite trihydrate and hexahydrate, and to establish the drying rate at various temperatures, so that dryer operation could be improved. The tests were performed in an oven operating at 40.5 kPa (12 in. Hg) vacuum and at 80°C (176°F), 120°C (248°F), and 150°C (302°F). At 80°C and 120°C, reaction rates were very slow for hexahydrate and nonexistent for trihydrate. At 150°C, hexahydrate dehydrated to trihydrate in two hours, and then gradually lost water until it reached $MgSO_3 \cdot 1/2H_2O$ (apparent composition). At 150°C, the trihydrate was stable for 48 hours and then rapidly dehydrated to the apparent $MgSO_3 \cdot 1/2H_2O$. Additional work at 166°C (330°F) showed dehydration of the hexahydrate in 10 minutes, and of the trihydrate in 25 minutes. The dehydration study indicated that water molecules are more strongly bound in the trihydrate than in the hexahydrate.[117]

A comparison of the dryer product temperature with the percentage of dry solids ($MgSO_3$), as a function of the dryer gas outlet temperature, showed that significantly more heat energy must be used to dry the trihydrate, the predominant product at Boston Edison, than is needed to dry the hexahydrate, the predominant product at PEPCO (Figure 3-63).[118]

The SO_2-removal capability of regenerated MgO was also investigated. The size range of MgO particles returned for slaking was found to be very important in this respect. Experimental data indicate that grinding calcined magnesia increases its reactivity, but only within certain limits.[119] Laboratory data now indicate that a size range of between 75 and 150 μm (2.9×10^{-3} and 5.9×10^{-3} in.) is best for recycled, regenerated magnesia. The greatest increase in MgO activity occurs when particles of 295+ μm are reduced to 150 to 295 μm (5.9×10^{-3} to 1.2×10^{-2} in.) additional improvement in SO_2 removal is obtained when size is reduced to 75 to 150 μm (2.9×10^{-3} to 5.9×10^{-3}in.). Further reduction has only a slight effect on activity increase.

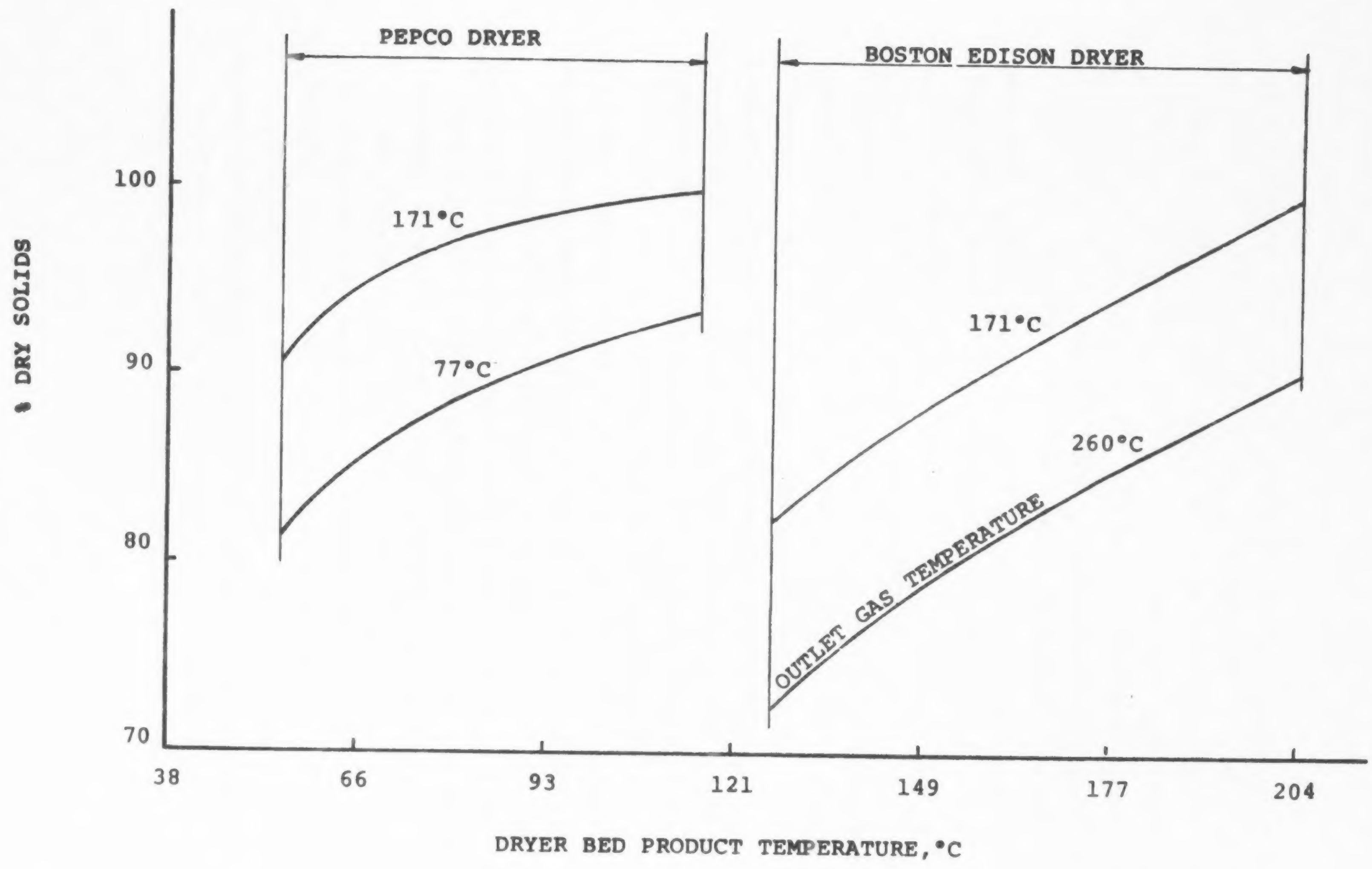


Figure 3-63. Comparison of PEPCO and Boston Edison Dryer Operation.

(For ranges of dryer gas outlet temperature)

The properties of regenerated MgO are determined by calcining conditions, as well as by $MgSO_4$ and fly ash levels.[120]

At PEPCO, flue gas was saturated and cooled in the wet particulate scrubber to 49° to 55°C conditions that favor hexahydrate formation in the absorber. At Boston Edison, flue gas entered the single-stage venturi scrubber at 138° to 160°C (280 to 320°F); favoring the formation of trihydrate.

Further work just completed by Radian Corporation determined the following:

1. The trihydrate is the thermodynamically stable form at scrubbing process conditions. Precipitation of the hexahydrate occurs as the result of kinetic phenomena. Precipitation may take place by two mechanisms: by nucleation, and by growth on existing crystals. Both mechanisms were studied. Nucleation and crystal growth rates of the hexahydrate are much faster than those of the trihydrate at conditions encountered in the magnesium oxide wet scrubbing process.[121]

2. Operating experience with magnesium oxide scrubbing processes indicates that magnesium sulfite hexahydrate forms under conditions where, theoretically, trihydrate is the thermodynamically stable form. Formation of the metastable hexahydrate is due to kinetic phenomena, which are described in this work.[122]

3. The system under investigation is complex and several phenomena can occur simultaneously. These include trihydrate nucleation, hexahydrate nucleation, trihydrate crystal growth, hexahydrate crystal growth, and hexahydrate dissolution.[123] The results obtained in scrubber-like systems were consistent with the results for tests of the individual phenomena.[124]

4. The following observations about the relative rates of the individual processes are important for design considerations. The rate of trihydrate

nucleation at magnesium oxide operating temperatures of 55°C (131°F) appears to be very slow compared with the rate of hexahydrate nucleations. Although hexahydrate nucleation was observed, evidence of trihydrate nucleation in scrubber-like media was not obtained. In addition, the rate of hexahydrate crystal growth was much faster than the rate of trihydrate crystal growth.[125]

5. The following sequence was observed in scrubber-like systems and the times were characterized: hexahydrate precipitation (tens of minutes), simultaneous hexahydrate dissolution and trihydrate precipitation (hundreds of minutes), trihydrate precipitation and attainment of equilibrium (thousands of minutes).[126]

6. By specifying the composition and particle size of seed crystals, the solution composition (driving force), and the reactor volume (residence time), it is possible to design a system to produce either of these hydrates as the solid product.[127]

7. Design tools used in other SO_2 slurry-scrubbing processes, such as the particle balance concept, are applicable to this system.[128]

8. Since nucleation rate is an important factor, equipment that cause high-energy changes in solution will have an effect on precipitation phenomena. Turbulence effects of nozzles, pumping, agitator speed, and control valve throttling will be important.[129]

9. Hold-tank and reactor volumes, and slurry density determine the solid residence times. Solid residence time in the system will be an important design criterion.[130]

10. Temperature and solution composition can be used to control seed crystal composition.[131]

11. In solutions containing only magnesium sulfite in water, the hexahydrate is the stable form at temperatures up to the transition temperature of 41°C (106°F). At higher temperatures, the tri-

hydrate is the form that precipitates from pure solutions. The analysis showed that the trihydrate is the thermodynamically stable species in solution compositions and at temperatures characteristic of the scrubbing process. Thus, the formation of hexahydrate crystals is explained on the basis of kinetic factors.[132]

3.4.5 Process Operability Considerations - Data on both systems indicate 27 to 80 percent operability, which is expressed as the ratio of scrubber operating time to boiler operating time for MgO systems applied to coal-fired boilers in the United States. When reviewing these operability levels, a number of factors must be kept in mind, including:

- Sulfur dioxide collection efficiency was frequently over 90 percent during test periods.
- Two units (Mystic and Dickerson) were trial installations, built to obtain operating data. As such, various construction materials were used that would not have been used in a long-term installation on a new plant.
- Handling and drying properties of the wet $MgSO_3$ were not adequately considered in system designs. This caused problems in the ancillary equipment serving the scrubber system, not in the scrubber itself.
- The sulfuric acid plant which was to receive SO_2 from the Eddystone MgO regeneration facility was shut down by its owner and a new site had to be found.
- The single regeneration facility at Rumsford, Rhode Island, could not simultaneously process material from Mystic and Dickerson.
- Many of the problems at Eddystone were related to particulate scrubbing and not to the SO_2 absorber section.
- Many problems encountered and solved during these early programs would not be repeated in subsequent designs.

This type of FGD system also has the distinct advantages of having no sludge disposal problem, and, thanks to its process chemistry, only minimal scaling and plugging problems.

3.4.5.1 Operational Problems and Solutions - Operational difficulties with the two trial systems have led to better design. The problems were primarily mechanical, caused by faulty equipment or by a lack of understanding of the nature of magnesium oxide and its sulfur compounds.

Heated makeup mixing tanks or slakers equipped with agitation and steam spargers are necessary if the production of makeup magnesium slurry is to be reliable. At Mystic and Dickerson, attempts to add dry magnesia without prior slurrying failed. Installation of a heated slurry tank with agitation solved this problem.[133,134,135] When regenerated mangesia was first used, it was found to be less reactive than virgin material; additional heat was required to slake it.[136,137] Other methods of introducing and slurrying magnesia were either unsuccessful, or less successful than high-speed agitation and steam sparging.[138] Heating the magnesia solution above 82°C (180°F) in the slaker must be avoided, since scaling will result in the tank and pipes.

Corrosion and erosion occurred in both the particulate scrubbing and the SO_2 absorbing systems, although rubber-lined pipes, pumps, and valves reduced the problem. Critical pumps should have installed spares.[139,140,141,142,143,144] The pH in the particulate scrubbing system had been reported[+] as low as 0.5.[145] This, coupled with the abrasiveness of the fly ash resulted in the erosion, corrosion and rapid failure of carbon steel. In the sulfur dioxide scrubbing system, the failure of carbon steel was due to the abrasiveness of magnesium oxide and to periodic pH swings from

---

[+] At the Eddystone Station.

0.5 to 9.0, caused by fluctuations in the magnesium oxide makeup slurry feed.[146] Inside the scrubbers, 316L stainless steel, Inconel 625, flaked glass, and corrosion-resistant polyester resin coatings gave the best service.[147,148,149] At Mystic and Dickerson, flaked glass and polyester resin provided excellent corrosion protection. There was no problem with scaling or buildup inside the scrubbers.[150,151,152,153]

Both magnesium sulfite trihydrate (small, hard-to-handle crystals) and hexahydrate (large, easy-to-handle crystals) are produced in the scrubber. The systems should be capable of handling the more troublesome magnesium sulfite trihydrate.[154,155,156]

Trihydrate requires half as much heat of vaporization to dewater the crystal as does hexahydrate; the latter loses water at 100° to 120°C (212 to 248°F), however, while trihydrate loses it at 180° to 190°C (356 to 374°F). Drying the hexahydrate may be accomplished with low pressure steam, while drying the trihydrate would probably require oil-fired heating.

Centrifuge bowl plows, head plows, and conveyors should all have hard surfaces (stellite, possibly ceramics) to reduce or eliminate the kind of wear problems that were experienced at Mystic and Dickerson.[157] Also, wash-out connections capable of thoroughly washing the centrifuge internals should be installed.[158,159,160] A final check on the stellite-hardened surfaces at Mystic indicated a life expectancy of over a year. The need for the wash-out connection was shown at both Mystic and Dickerson, where the centrifuges had to be dismantled to break loose hardened material. Seizing of the centrifuge internals ended the Dickerson trial prematurely.[161] Consequently, in keeping with the common practice, spare centrifuge should be included in any new unit.[162,163,164]

Further, the centrifuge should be designed to handle a wide range of concentrations and viscosities of solids, while giving good separation. This is necessary because higher-than-normal solids levels, the result of excessive magnesium sulfite trihydrate concentrations and higher-than-normal viscosities, have caused unit shutdown.[165,166,167,168]

The centrifuge discharge hopper must be designed to prevent the product from building up. Downstream surges, caused by sudden release of large masses of wet magnesium sulfite cake, have caused equipment overloads, particularly in conveyors and elevators.[169,170]

The magnesium sulfite dryer should be cocurrent and have internal chain sections and external rappers to dislodge built-up product.[171] Excessive sulfite carryover from the dryer at Mystic illustrated the dangers of not having cocurrent flow.[172,173,174]

The bucket elevator for transporting dried magnesium sulfite to storage should have built-in excess capacity, so that a product surge will not cause it to overload. The elevator proved to be a bottleneck during prototype operations, causing the whole system to operate at less than capacity.[175,176]

The dryer off-gas system should include a cyclone separator for removal of entrained magnesium sulfite; the "cleaned" gas stream should then be vented into the sulfur dioxide scrubber for further magnesium sulfite removal. Finally, sulfite collected in the cyclone separator should be discharged through a rotary valve into a pneumatic conveying system and taken to storage.[177]

3.4.5.2 Technological Transfer from the Pulp and Paper Industry - Magnesium-based pulping operations have been in operation in the pulp and paper industry since the 1940's. The magnesium process (known as Magnefite) is the most amenable to regeneration of the four primary pulping systems (magnesium, calcium, ammonium, and sodium). As a rule waste gas, 1.0 to 1.5 percent of which is SO_2, is precooled to 43°C (110°F) to 66°C (150°F) prior to SO_2 absorption. The scrubbing slurry consists of $Mg(OH)_2$ and $MgSO_3$ in water. The exit gas normally contains less than 250 ppm SO_2, representing 97.5 to 98 percent SO_2 removal; and overall operability is high. The $MgSO_3$ calcining takes place in a reducing atmosphere. The regenerated MgO is slaked at 83°C (180°F) to 94°C (200°F) prior to addition as makeup to the absorber slurry.[178]

3.4.5.3 Product Disposal - The magnesium oxide process is a regenerable process, and as such does not present the landfill and disposal problems associated with the "throw-away" sulfur dioxide scrubbing processes (lime, limestone, or double alkali). The magnesium sulfite produced, however, must be regenerated. Points that must be considered in the operation of an MgO system include the proximity of the unit to a regeneration facility (if off-site); the freight costs involved in transporting magnesium sulfite to that facility; and the market for the resulting sulfuric acid (although the sulfur dioxide, sulfur trioxide, and sulfur are other possible products).

3.4.6 Pilot Plant and Prototype Potential

Babcock and Wilcox (B&W) carried out a test program for the EPA, developing the scrubbing step of the MgO FGD process. They used a TCA absorber, which was later adapted for use at the successful MgO FGD system at Onahama Smelting and Refining.

The two Chemico MgO FGD systems (Oil-fired at Mystic; Coal-fired at Dickerson) were prototype units, built to demonstrate the potentials of the process and to determine the major areas for improvement. The venturi was adopted for use in these prototypes. The units were built on a low budget and include little redundancy, to enable a high percentage of on-stream time. Considering this, they achieved their purpose.

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3.5 THE WELLMAN-LORD PROCESS

This scrubbing system uses an aqueous sodium sulfite solution to absorb SO_2. Sodium bisulfite is formed, and the SO_2 is released in a forced-circulation evaporator-crystallizer. The regenerated sodium sulfite is crystallized and dissolved for recycle to the absorber. The water vapor in the concentrated SO_2 stream is condensed, and the resulting stream is recovered as liquid SO_2, sulfuric acid, or elemental sulfur. This system is used on combustion and other SO_2 emission sources.

A typical Wellman-Lord system as applied to a combustion process is shown in Figure 3-64.

3.5.1 Process Description[1,2]

Primary removal of fly ash and other particulates is generally provided by an ESP prior to the flue gas stream entering the SO_2 absorber. A variable-throat venturi prescrubber provides cooling of flue gases to 49 to 54°C (120 to 130°F), additional particulate removal, and primary chloride and sulfur trioxide removal. The need for an ESP may be avoided by use of a larger, more efficient venturi; however, the process vendor, Davy Powergas, does not recommend this because of the possibility of reduced mechanical reliability.[3,4]

This recovery process is based on the chemistry of the sodium sulfite/bisulfite system. After appropriate pretreatment, the flue gas containing SO_2 enters the absorber where the SO_2 is contacted countercurrently in the absorber by a sodium sulfite solution. As the sodium sulfite absorbs and reacts chemically with the SO_2, forming the more-soluble sodium bisulfite, the solution becomes less saturated. This means that fouling and scaling are unlikely to occur in the absorber.[5]

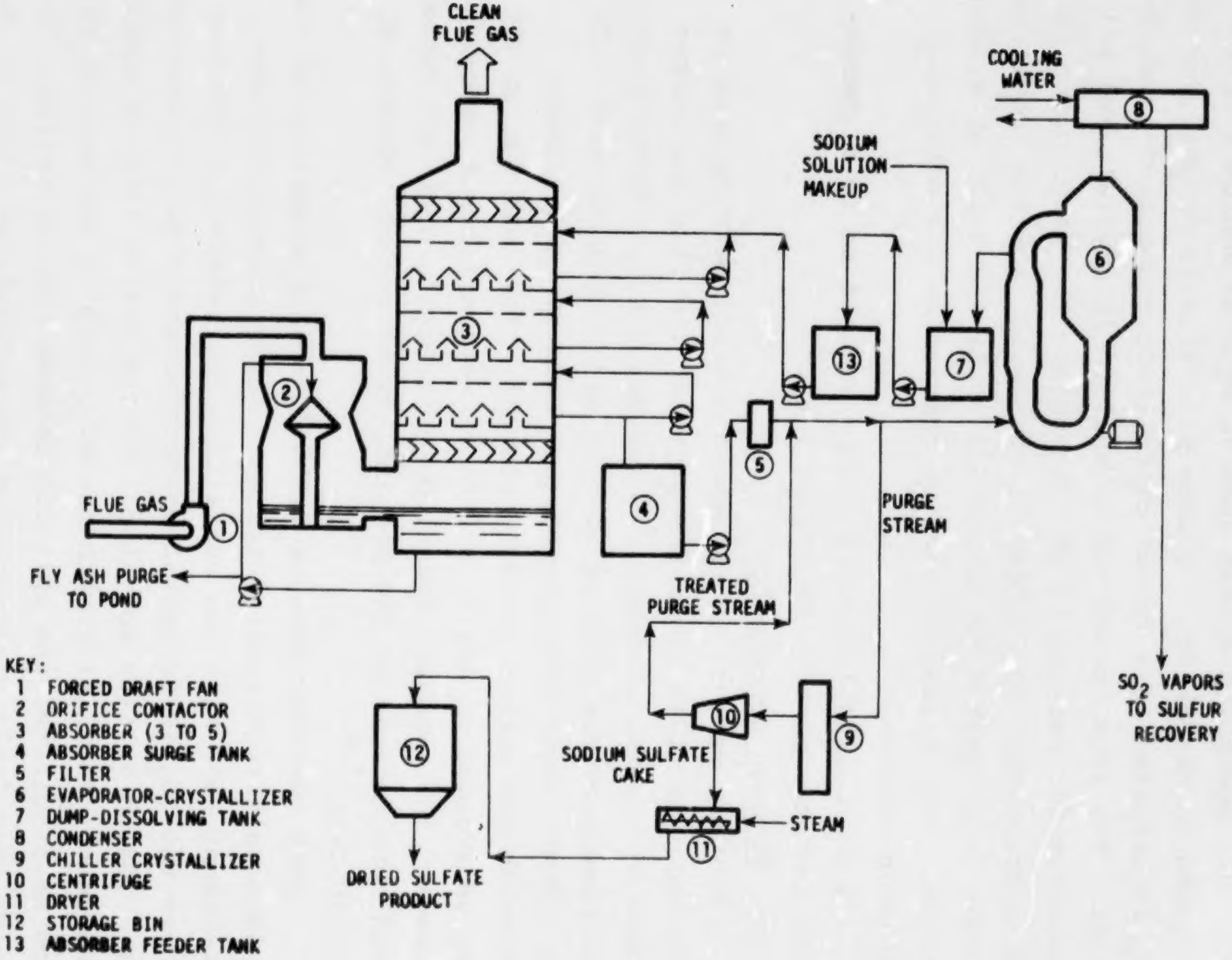

Figure 3-64. Typical Wellman-Lord SO_2 Recovery Process.

Oxygen and sulfur trioxide in the flue gas react with the sodium sulfite, forming the unreactive sodium sulfate/ bisulfate. The presence of the unreactive species requires a purge from the absorber to maintain the level of reactive sodium sulfite and to reduce the possibility of scaling problems.[6,7]

The absorber tower has a series (typically three to five) of valve and collector trays made of Hastalloy G (bottom trays) or 316L stainless steel. Because of the extremely low L/G ratio 0.8 to 1.35 l/m^3 (6 to 10 gal/1000 acf) required for absorption, each set of trays requires recirculation of the sodium sulfite absorbing solution. This recirculation provides the necessary amount of liquid required to load the valve trays hydraulically.[8] The sodium bisulfite-rich solution is discharged from the bottom of the absorber into a surge tank. It is then pumped to the sulfur recovery area for regeneration.

A mist eliminator removes entrained liquid droplets from the cleaned flue gas stream. The gases then enter a reheater to eliminate the steam plume, to provide buoyancy to the exhaust gas, and to minimize corrosion of the stack from condensate.[9]

The chemical reactions in the SO_2 absorber are as follows:

SO_2 Absorption - Sulfur dioxide and sodium sulfite react to form sodium bisulfite.

$$SO_2 + Na_2SO_3 + H_2O \rightarrow 2\ NaHSO_3 \qquad (1)$$

Oxidation - Some oxidation of sodium sulfite to sodium sulfate occurs.

$$2Na_2SO_3 + O_2 \rightarrow 2Na_2SO_4 \qquad (2)$$

The sodium ion makeup reactions are as follows:

Sodium carbonate (soda ash) or sodium hydroxide (caustic) reacts with sodium bisulfite to regenerate the SO_2 absorbent, sodium sulfite.

$$Na_2CO_3 + 2NaHSO_3 \rightarrow 2Na_2SO_3 + H_2O + CO_2\uparrow \qquad (3)$$

$$NaOH + NaHSO_3 \rightarrow Na_2SO_3 + H_2O \qquad (4)$$

The product solution from the absorber is pumped through filters for removal of any fly ash or other particulates. A small side stream of the filtered solution is pumped to the purge treatment area for sodium sulfate removal. The bulk of the stream is pumped to the evaporator for regeneration as sodium sulfite.

Thermal regeneration of the bisulfite-rich absorbing solution is accomplished in a forced circulation vacuum evaporator. The sodium bisulfite in the feed solution is indirectly heated by low pressure steam and converted to sodium sulfite by driving off SO_2 and water vapor. The sodium salts remaining in the evaporator circulating liquid now form a crystalline slurry, the major component of which is sodium sulfite. The concentration of crystals in this evaporator circulation slurry is kept at a predetermined level to minimize heat transfer area fouling. The slurry is withdrawn continuously to the dissolving tank, where the condensate from the evaporator is used to dissolve the crystals. The resulting solution is mixed with makeup sodium solution and pumped back to the top tray of the absorber.[10,11,12,13,14]

The overhead SO_2-water vapor mixture is concentrated through a series of partial condensers. Depending upon the size of the chemical plant, which is based on the amount of sulfur dioxide absorbed, the first partial condenser for a portion of the SO_2-water vapor may be the heat exchanger for a second stage evaporator. As the size of the chemical plant increases somewhere beyond the production of 2268 kg (5000 lbs) of SO_2 per hour, operating economics favor a double-effect evaporator. A double-effect evaporator may reduce steam consumption by 40 to 45 percent. The condensers following the evaporators may be air- or water-cooled, depending on the process absolute pressure, product purity required, and cooling medium available. The product SO_2, at least in the case of a power plant, will be used to make elemental sulfur or sulfuric acid; it is therefore advantageous to have the SO_2 concentration as high as possible. Normally, this product gas will be approximately 85-volume-percent SO_2, with the balance being water vapor.[15,16]

The chemical reactions in the regeneration plant are:[17,18,19]

Decomposition of the sodium bisulfite to sodium sulfite SO_2.

$$2NaHSO_3 \xrightarrow[100^\circ C]{\Delta} Na_2SO_3 + H_2O\uparrow + SO_2\uparrow \tag{5}$$

A small amount of the sodium bisulfite may react to form other species when heated as follows:

Sodium pyrosulfite ($Na_2SO_2O_5$)

$$2NaHSO_3 \xrightarrow{\Delta} Na_2S_2O_5 + H_2O\uparrow \tag{6}$$

Sodium thiosulfate ($Na_2S_2O_3$)

$$\text{Reaction unclear} \xrightarrow{\Delta} Na_2S_2O_3 \tag{7}$$

Sodium dithionate ($Na_2S_4O_6$)

$$\text{Reaction unclear} \xrightarrow{\Delta} Na_2S_4O_6 \tag{8}$$

The sodium thiosulfate may be in the 0.2- to 0.3-percent range. The other side reaction products are normally present only in trace amounts.[20]

The resulting concentrated (85 percent) SO_2 stream can be converted to elemental sulfur, sulfuric acid, or liquid SO_2. The reactions are as follows:[21,22]

Sulfuric Acid - Sulfur dioxide reacts with oxygen in the presence of vanadium pentoxide catalyst to form sulfur trioxide

$$2SO_2 + O_2 \rightarrow 2SO_3 \qquad (9)$$

The sulfur trioxide reacts with water to form sulfuric acid.

$$SO_3 + H_2O \rightarrow H_2SO_4 \qquad (10)$$

Sulfur - Methane (natural gas) reacts with sulfur dioxide to form hydrogen sulfide.

$$2CH_4 + 3SO_2 \rightarrow 2CO_2 + 2H_2O + 2H_2S + S \qquad (11)$$

The hydrogen sulfide reacts with more sulfur dioxide, forming water vapor and sulfur.

$$2H_2S + SO_2 \rightarrow 2H_2O + 3S \qquad (12)$$

The overall reaction is:

$$CH_4 + 2SO_2 \rightarrow CO_2 + 2H_2O + 2S \qquad (13)$$

Other commercially available reductants that can be used in place of methane (natural gas) are carbon monoxide, hydrogen, higher hydrocarbons up through propane, and the products of coal gasification (low-, medium-, and high-Btu coal gases).

Liquid SO_2 - The sulfur dioxide vapors are dried with silica gel to remove moisture. The vapors then are compressed and condensed; the resulting liquid is collected and stored in pressurized tanks.

The purge streams go to a chiller crystallizer. With controlled crystallization, the sulfate precipitates in a much greater proportion than the other sodium compounds. Thus, the solid phase has a high sulfate concentration while the liquid phase has a higher sulfite concentration. The solid phase is removed in a centrifuge and the liquid is recycled to the process. The sulfate can be dried for sale or disposal, or can be neutralized and discharged as an innocuous effluent.[23]

Table 3-43 represents a material balance for the Wellman-Lord system as applied to a 500-MW, 3.5 percent-sulfur coal-fired plant. The flow streams refer to Figure 3-65.

3.5.2 Wellman-Lord FGD Systems in the United States

Seven Wellman-Lord systems are currently in operation in the United States as listed in Table 3-44. Six of these units are installed on sulfuric acid or Claus sulfur recovery units. The gas flow rates on these are small, 51,000 to 133,000 nm^3/hr (30,000 to 78,000 scfm), in comparison with a new 500-MW coal-fired boiler with a gas flow rate of 1,700,000 nm^3/hr (1,000,000 scfm). The inlet SO_2 concentrations on the seven small units range from 2700 to 10,000 ppm, which is about one to five times the SO_2 concentration expected with a 3.5-percent-sulfur coal-fired boiler. The smaller units, however, operate on fairly clean, dry streams with low oxygen concentrations in comparison with boiler flue gases.[25,26]

The SO_2 removal efficiency of the six units is typically 90 percent or greater; removal efficiencies in excess of 97 percent are reported. The collected SO_2 is either recycled to the sulfuric acid or Claus sulfur unit. Little operational data are available; it is reported, however, that the six units have absorber on-stream times of greater than 97 percent.[27,28]

The No. 11 unit at the D.H. Mitchell Generating Station of Northern Indiana Public Service Company is currently the only utility operation in the United States, and also it is the only coal-fired application.[29] It is described further in the next Section. In addition, three Wellman-Lord systems

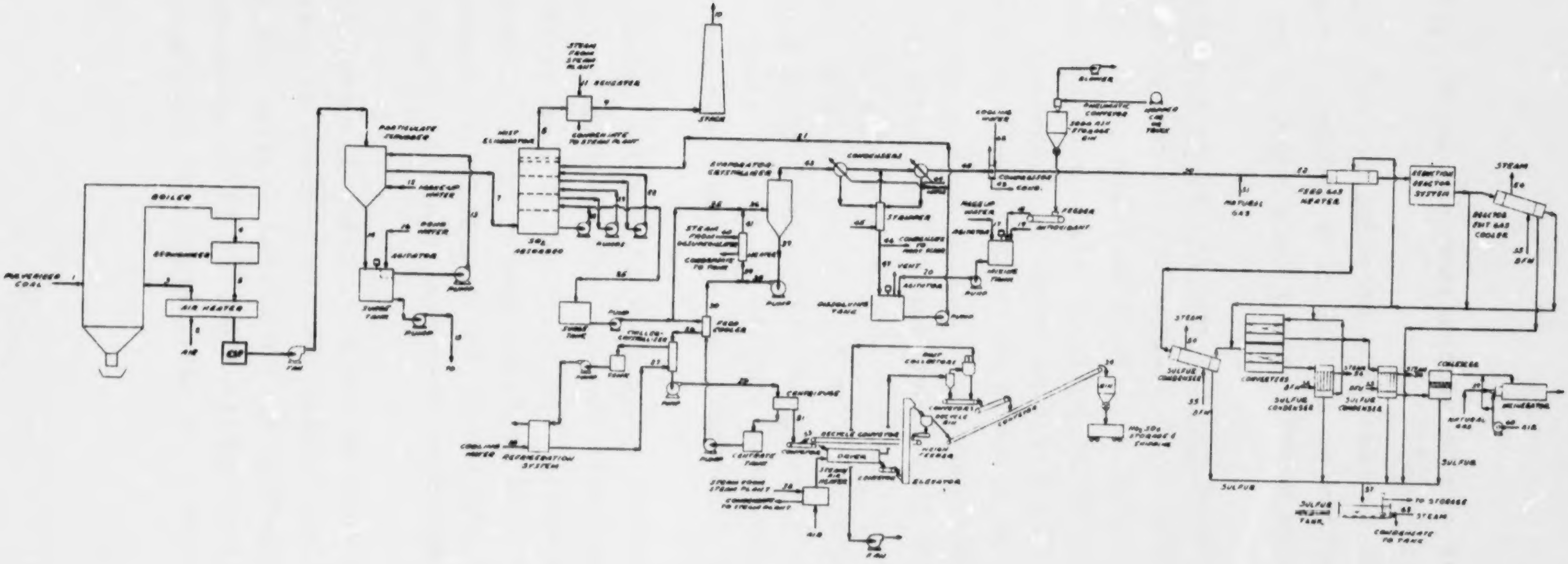

Figure 3-65. Flow diagram of the Wellman-Lord SO_2 recovery process and a sulfur removal unit.

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Table 3-43. MATERIAL BALANCE FOR A WELLMAN-LORD FGD SYSTEM SERVING A 500-MW BOILER

Notes: 1. M indicates 1,000
2. Calculations based upon the following:
  a. 105 percent stoichiometric sodium
  b. 9,000 Btu/kWh for the unit
  c. 12,000 Btu/pound of coal
  d. 3.5 percent sulfur coal (dry basis)
  e. 92 percent of the sulfur in the coal evolves as SO_2
  f. 14 percent ash in the coal (as fired basis)
  g. 75 percent of the ash in the coal evolves as fly ash
  h. 99.0 percent removal of particulates by the ESP

Stream No.	1	2	3	4
Description	Coal to the boiler	Combustion air to air heater	Combustion air to the boiler	Gas to the economizer
Rate, lb/hr	375 M	4,518 M	4,518 M	4,518 M
scfm		984 M	984 M	984 M
gpm				
Particulates, lb/hr				39.3 M
Temperature, °F		110	535	890
Specific gravity				
Viscosity, cps				
Undissolved solids, %				
pH				

Table 3-43 (Continued).

Stream No.	5	6	7	8	9	11
Description	Gas to the ESP	Gas to scrubber	Gas to SO_2 absorber	Gas to reheater	Gas to stack	Steam to gas reheater
Rate, lb/hr	4,518 M	4,518 M	4,518 M	4,518 M	4,518 M	15 M
scfm	984 M	984 M	984 M	984 M	984 M	
gpm						
Particulates, lb/hr	39.3 M	393				
Temperature, °F	310	310	127	127	170	470
Specific gravity						
Viscosity, cps						
Undissolved solids, %						
pH						

Stream No.	12	13	14	15	16
Description	Make up water to scrubber	Recycle slurry to scrubber	Scrubber slurry to surge tank	Particulate slurry purge	Pond water recycle
Rate, lb/hr	202 M	9,764 M	9,796 M	224 M	190 M
scfm					
gpm	403	19.0 M	19.0 M	410	381
Particulates, lbs/hr					
Temperature, °F			127		
Specific gravity		1.03		1.09	
Viscosity, cps					
Undissolved solids, %		5		15	
pH			3		

Table 3-43 (Continued).

Stream No.	17	18	20	21	22
Description	Make up water to mix tank	Soda ash to mix tank	Make up to dissolving tank	Feed solution to SO_2 absorber	Recycle to third stage
Rate, lbs/hr	10.6 M	2,650	13.3 M	349 M	2,052 M
scfm					
gpm	21.2		22.2	564	3260
Particulates, lb/hr					
Temperature, °F			100		
Specific gravity			1.20		
Viscosity, cps					
Undissolved solids, %					
pH					

Stream No.	23	24	25	26	27	28
Description	Recycle to second stage	Recycle to first stage and product	Liquor to surge tank	Liquor to chiller	Solution to chiller	Cooling water-refrigeration
Rate, lb/hr	2,432 M	2,820 M	370 M	214 M	1,395 M	389 M
scfm						
gpm	3,828	4,408	580	334	2,307	777
Particulates, lb/hr						
Temperature, °F				68	25	
Specific gravity					1.07	
Viscosity, cps						
Undissolved solids, %						
pH						

Table 3-43 (Continued).

Stream No.	29	30	31	32	34	35	36
Description	Slurry centrifuge	Centrate to heater	Cake to conveyor	Steam to heater	Sulfate to bin	Liquor to evaporator	Feed to evaporator
Rate, lb/hr	214 M	205 M	9,380	20.4 M	3,712	156 M	120 MM
scfm							
gpm	334	320				244	137 M
Particulates, lb/hr							
Temperature, °F	35	98		470			
Specific gravity							
Viscosity, cps							
Undissolved solids, %							
pH							

Stream No.	37	38	39	40	41	43	44
Description	Slurry from evaporator	Slurry to heater	Feed to heater	Steam to heater	Slurry evaporator recycle	Vapor to primary condenser	Water to condensers
Rate, lb/hr	120 MM	120 MM	120 MM	230 M	120 MM	216 M	11,136 M
scfm						70.8 M	
gpm	137 M	137 M	137 M		137 M		22.2 M
Particulates, lb/hr							
Temperature, °F				220			
Specific gravity							
Viscosity, cps							
Undissolved solids, %							
pH							

Table 3-43 (Continued).

Stream No.	45	46	47	48	49	50
Description	Steam to stripper	Condensate to scrubber	Condenate to dissolving tank	Gas to compressor	Condensate	Gas to SO_2 reduction
Rate, lb/hr	7,336	10,326	191 M	21.4 M	738	20.6 M
gpm		20.6	382			
Particulates, lb/hr						
Temperature, °F	220					250
Specific gravity						
Viscosity, cps						
Undissolved solids, %						
pH						

Stream No.	51	52	53	54	55	57	59
Description	CH_4 to SO_2 reduction	Gases to reheater	Recycle to dryer	Steam to steam plant	Boiler feed water to condensers	Sulfur to storage	CH_4 to incinerator
Rate, lb/hr	3,306	23.9 M	55.5 M	9.5 M	13.7	9,342	243
scfm	1,130	3,410					83
gpm					116	10.4	
Particulates, lb/hr							
Temperature, °F	200		220	250	230		
Specific gravity							
Viscosity, cps							
Undissolved solids, %							
pH							

Table 3-43 (Continued).

Stream No.	60	62	63
Description	Air to incinerator	Water to compressor	Steam to sulfur storage
Rate, lbs/hr	8,813	148 M	3000
scfm	1,930		
gpm		296	
Particulates, lb/hr			
Temperature, °F	70		290

Table 3-44. OPERATIONAL WELLMAN-LORD FGD SYSTEMS IN THE U.S.[24]

Company/location	Completion date	Feed gas origin	Gas flow, 1000 Nm^3/hr	Gas flow, (scfm)	Design SO_2 concentration, ppm	Disposition of SO_2
Olin Chemical* Paulsboro, New Jersey	July 1970	Sulfuric acid plant	76	(45,000)	in 6,000 out 500	Recycle to acid plant
Std. Oil of California El Segundo, California	September 1972	Claus plants	51	(30,240)	in 10,000 out 250	Recycle to Claus plants
Allied Chemical Calumet, Illinois	November 1972	Sulfuric acid plants	51	(29,850)	in 2,700 out 250	Recycle to acid plants
Olin Chemical Curtis Bay, Maryland	May 1973	Sulfuric acid plants	133	(78,046)	in 4,000 out 250	Recycle to acid plants
Std. Oil of California Richmond, California	August 1974	Claus plant	51	(30,000)	in 10,000 out 250	Recycle to Claus plant
Std. Oil of California El Segundo, California	January 1975	Claus plant	51	(30,000)	in 10,000 out 250	Recycle to Claus plant
Northern Indiana Public Service Gary, Indiana	December 1976**	Coal-fired boiler (115 MW)	527	(310,000)	in 2,200 out 200	Elemental sulfur

* Plant operation suspended as of January 1, 1976.

**First integrated operation of the plant was in December, 1976; however, the Wellman-Lord system began operations in July 1976, before completion of the Allied sulfur recovery unit.

are currently under construction, (Table 3-45).[30,31]

The largest Wellman-Lord installations are being constructed on four coal-fired boilers at Public Service Company of New Mexico's San Juan Station, Fruitland, New Mexico. The retrofit installations on Units 1 and 2 should start-up in early 1978; however, an explosion in Boiler 2 has delayed start up of that unit until mid-1978. The installation on the new No. 3 boiler is expected to be completed in 1979. Table 3-46 provides further details on these units. The coal that is burned has the following specifications: sulfur - 1.3 percent; ash - 22.4 percent; moisture - 14.8 percent; and heating value - 18,870 kJ/kg (8100 Btu/lb). On all three units, the Wellman-Lord system is preceded by an ESP. Because of the size of the units, multiple absorber modules are being installed. Rectangular absorbers are used with tile-lined concrete walls that are resistant to corrosion from the circulating solution. The systems are designed for 90-percent SO_2 removal, or better and elemental sulfur will be recovered from the SO_2 collected. The prescrubber blowdown will be used to condition the fly ash for hauling by truck and disposal by burial in an adjacent coal mine area.[34,35,36,37]

A more recent Wellman-Lord project is an installation for ARCO/Polymers at Monaca, Pennsylvania, where a single scrubber will receive the flue gas from three coal-fired boilers with a combined generating capacity of 100 MW.[39,40]

3.5.2.1 Northern Indiana Public Service Co. (NIPSCO) - D.H. Mitchell Unit No. 11 System Description - The 115-MW unit at NIPSCO is the first application of a Wellman-Lord system to a coal-fired utility boiler, (Figure 3-66). Flue gases from the boiler pass through an ESP for primary particulate removal (about 98.5%-efficiency). The booster blower delivers

Table 3-45. WELLMAN-LORD FGD SYSTEMS UNDER CONSTRUCTION IN THE UNITED STATES[32,33]

Company/location	Feed gas origin	Gas flow, 1000 Nm^3/hr	Gas flow, (scfm)	Design SO_2 concentration, ppm	Disposition of SO_2
Public Service Co. of New Mexico, San Juan Station Fruitland, New Mexico	Coal-fired boilers (700 MW total)	3058	(1,800,000)	in 850 out 85	Elemental suflur
Public Service Co. of New Mexico, San Juan Station Fruitland, New Mexico	Coal-fired boilers (184 MW total)	889	(523,000)	in 850 out 85	Elemental sulfur
ARCO/Polymers Monaca, Pennsylvania	Coal-fired boilers (100-MW)	425	(250,000)	in 2000 out 250	Unknown

Table 3-46. DESIGN PARAMETERS FOR WELLMAN-LORD FGD INSTALLATIONS AT SAN JUAN STATION OF PUBLIC SERVICE COMPANY OF NEW MEXICO[38]

	Unit 1	Unit 2	Unit 3	Unit 4
Boiler rating, MW (Gross)	350	350	550	550
Fuel rate, kg/hr	181,440	181,440	303,458	303,458
(lb/hr)	(400,000)	(400,000)	(669,000)	(669,000)
Gas flow, m^3/s scfm at 26.7°C	416.2	373.7	603.7	603.7
(80°F)	(882,000)	(792,000)	(1,279,400)	(1,279,400)
Inlet SO_2 concentration, ppm				
Minimum	490	490	490	490
Maximum	1200	1200	1200	1200
Average	850	850	850	850
Inlet particulate loading				
$kg/10^8$ J	0.0030	0.0043	0.0039	0.0039
$(lb/10^6$ Btu)	(0.07)	(0.10)	(0.09)	(0.09)
g/m^3	0.121	0.18	0.13	0.13
(gr/scf dry)	(0.053)	(0.079)	(0.056)	(0.056)
No. of scrubber trains[a]	4	4	5	5
m^3/s	207.2	200.1	240.8	240.8
(acfm/train)	(439,000)	(424,000)	(510,000)	(510,000)
Heating rate, kJ/kWh (Btu/kWh)				
Without scrubber	10,223 (9,682)	10,203 (9,663)	10,233 (9,691)	10,233 (9,691)
With scrubber	10,680 (10,115)	10,660 (10,096)	11,215 (10,621)	11,215 (10,621)
Sulfur regenerated, kg/hr (ton/hr)				
Design	2041 (2.25)	2041 (2.25)	2041 (2.25)	2041 (2.25)
Actual	852 (0.94)	852 (0.94)	1660 (1.83)	1660 (1.83)

[a] Each boiler will have an additional spare train to provide reserve capacity and improve reliability.

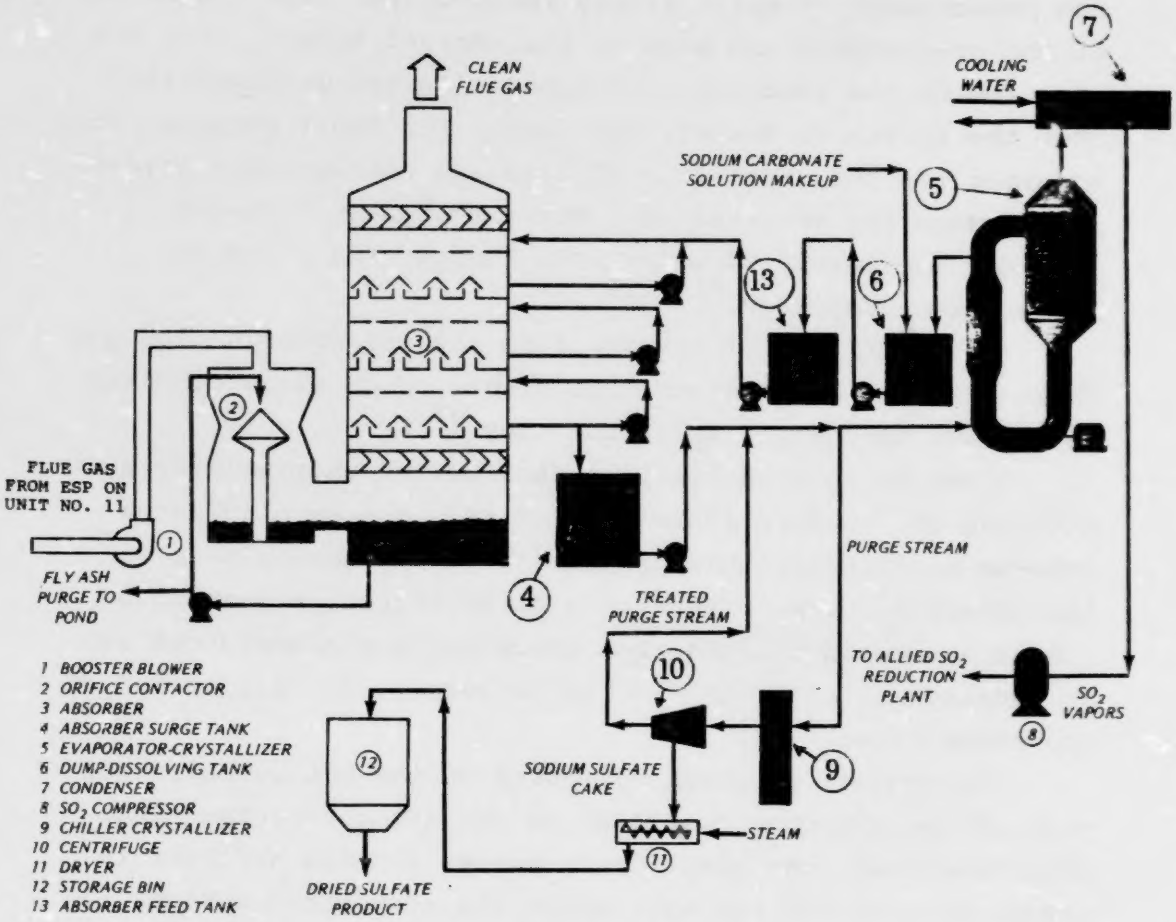


Figure 3-66. Schematic of the Wellman-Lord SO_2 - recovery process at NIPSCO.[46]

the flue gas through a variable-throat venturi prescrubber to the absorber. The flue gas is cooled and saturated in the prescrubber by water/slurry recirculated from the bottom of the prescrubber and back to the venturi sprays. Fly ash captured by the scrubbing solution is purged continuously from the system to the fly ash pond. For short periods, the scrubber has the capability of handling considerably greater than design fly ash loading. Water from Lake Michigan is used for the makeup of water lost through the purge and evaporation processes.[41,42,43,44]

The absorption of the SO_2 from the prescrubbed flue gas takes place in a three-stage absorber. Each stage consists of a valve tray and a collector tray.[45]

A sodium sulfite solution absorbs and chemically reacts with the SO_2 to form sodium bisulfite. A mist eliminator removes entrained liquid droplets from the gas exiting via the absorber stack. There is a direct-fired, natural gas reheat system in the absorber stack so that cleaned gas can be reheated to 82°C (180°F), if necessary, for dispersion of the steam plume.[47,48]

The product solution collected on the bottom collector tray of the absorber overflows to the absorber surge tank. From this tank, the solution is pumped through a filter to ensure that no fly ash will enter the evaporator system. A small sidestream of the filtered solution is sent to the purge treatment area to remove the sodium sulfate. The purge treatment equipment consists of four chilled-wall crystallizers; a slurry of sodium sulfate crystals forms in these crystallizers and is then removed in a centrifuge. The bulk of the product solution is pumped to the evaporator for regeneration of the sodium sulfite.[49,50]

The evaporation system consists of a forced-circulation vacuum evaporator. The filtered solution is recirculated in the evaporator, where low-pressure (345 kPa, 50 psig) steam is used to evaporate the water from the sodium bisulfite solution. When sufficient water has been removed, sodium sulfite crystals form and precipitate. Sulfur dioxide is removed with the overhead vapors. The slurry formed by the sodium sulfite crystals is withdrawn continuously to the dump/dissolving tank, where condensate from the evaporator is used to dissolve the crystals in the solution that is pumped back to the top stage of the absorber.[51,52]

Water vapor is removed from the SO_2 in water-cooled condensers. The SO_2 is compressed by a liquid ring compressor for introduction to the Allied Chemical SO_2-reduction facility. The gas stream is about 85 percent SO_2, the remainder mostly water vapor.[53,54]

Sodium lost as sulfate in the purge treatment system is replenished by the addition of sodium carbonate to the absorber solution. Soda ash is brought to the plant in trucks and transferred to the storage bin by a pneumatic conveying system. It is metered to the slurry tanks by a bin activator and belt feeder. The soda ash slurry is pumped to the absorber feed tank by parallel centrifugal pumps.[55,56]

The small sidestream of filtered solution from the absorber is pumped to four chilled-wall crystallizers, where sodium sulfate crystals form. The crystallized slurry is centrifuged to extract the sodium sulfate crystals, and clear solution is returned to the evaporator feed system. The sodium sulfate crystals are melted and fed to a steam-heated dryer, whose discharge product is then stored in a bin until loaded in trucks for shipment. Any gases that evolve from the purge treatment are chemically scrubbed and vented to the atmosphere.[57]

The compressed SO_2 is fed to the Allied Chemical SO_2-reduction plant, where it is reacted with natural gas. The resulting elemental sulfur is condensed and stored in molten form for shipment. The off-gases are burned in a tail gas incinerator to convert any hydrogen sulfite to SO_2 and returned to the absorber inlet.[58,59]

Distinguishing features - This unit is characterized by the following:

- Fired with 3.15 to 3.5-percent-sulfur coal.[60]
- Most of the fly ash is removed in the ESP (primary particulate removal) with some additional fly ash removal in the wet venturi prescrubber.[61,62]
- The absorber has three valve trays with separate recirculation pumps for each stage.[63]
- The SO_2 is recovered as elemental sulfur.[64,65]

Process guarantees - The contracts between NIPSCO and EPA and the other principals, Davy Powergas and Allied Chemical, contain penalty-assessable guarantees for the following:

° SO_2 emission levels - the process, according to the contracts, will be at least 90-percent efficient for SO_2 removal when firing coal with a sulfur content of up to 3.5 percent (approximately 2300 ppm by volume of SO_2 in the stack gas). As an override, no more than 200 ppm by volume of SO_2 will be present in the exit gas.[66,67,68]

° Mechanical reliability - The process plants will be mechanically sound.[69]

° Utilities consumption - The aggregate cost of steam, electric power, and natural gas required for operation of the complex will not exceed a specified cost; i.e., \$56.00 per hour based on:[70,71]

Electric power - \$0.007/kWh

Steam - \$0.50/453.6 kg (1000 lb) at 3792 kPa gauge (550 psig), 399°C (750°F)

Natural gas - \$0.55/1056 million J (million Btu)

° Chemical requirement - The quantity of makeup chemicals (caustic soda or soda ash) will not exceed a fixed daily amount. The average chemical makeup over a 12-day operating period at an average of 92 MW shall be no greater than 6.6 ton of Na_2CO_3/day.[72,73]

° Product quality - The sulfur product is to be at least 99.5-percent pure sulfur, and of a quality suitable for use in the manufacture of sulfuric acid by the contact process.[74,75]

Operating history - Initial start-up of the NIPSCO Wellman-Lord absorber was on July 19, 1976. An extended shutdown period began on November 28, 1976, when high-pressure steam supply failures from the boiler and from emergency back-up systems to the FGD plant resulted in freeze damage to the FGD plant. The shutdown lasted until early January 1977. During the period from July through November 1976, the Unit 11 boiler operated for 121 full days and 10 partial days, while the SO_2 removal system of the FGD plant operated for 71 full days and 23 partial days and was down for 38 days. The steam supply failures previously mentioned were responsible for 28 of the 38 days that the system was down.[76]

During the three sustained operating periods (Table 3-47), the absorber demonstrated the capability of greater SO_2 removal than specified in the performance criteria. The concentrations are shown in Figures 3-67, 3-68, and 3-69.[77]

Table 3-47. NIPSCO OPERABILITY DATA[78,79]

Run number	Duration (days)	Period
1	15	Sept. 25 through Oct. 9, 1976
2	11	Oct. 13 through Oct. 23, 1976
3	14	Nov. 15 through Nov. 28, 1976
Acceptance	16 1/2	Aug. 29 through Sept. 14, 1977 [12 days at 151.0 actual m^3/s (320,000 acfm) (92 MW equivalent) and 83 hours at 18.40 actual m^3/s (390,000 acfm) (110 MW equivalent)]

During runs 1, 2, and 3, the booster fan delivered the flue gas at the fixed rate of 544,000 actual nm^3/hr (320,000 acfm), the 92-MW design level, to attain the acceptance test levels. The load on Unit 11 varied from 60 MW to 108 MW. For safety reasons the multileaf stack damper was open during the operating periods; this allowed additional flue gas with lower SO_2 concentrations from Unit 6 to be pulled across the stack to the absorber.[83]

A boiler-related mishap on Jan. 15, 1977 caused Unit 11 to be shut down for repairs until May, 1977. The Wellman-Lord unit was placed in service again on June 13, 1977. The acceptance trail began on August 29, 1977, and was successfully completed on September 14, 1977. The Wellman-Lord FGD system was officially accepted from Davy Powergas by NIPSCO on September 16, 1977.

The 1-year demonstration run commenced September 15, 1977. During the trial, the evaporator and the absorber both operated without interruption, except for two periods (19 hours total) when the blower booster fan was shut down

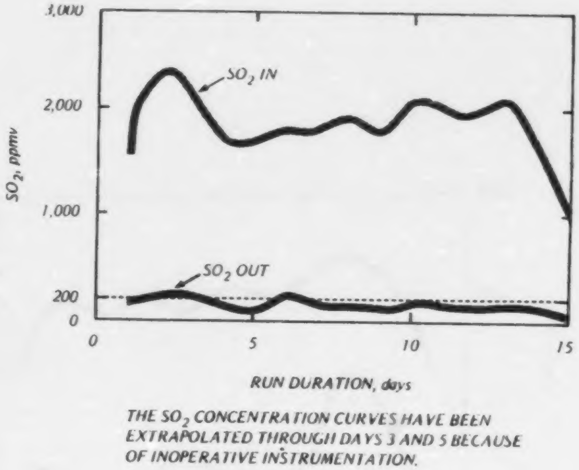


Figure 3-67. Inlet and outlet SO_2 concentrations during run no. 1.[80]

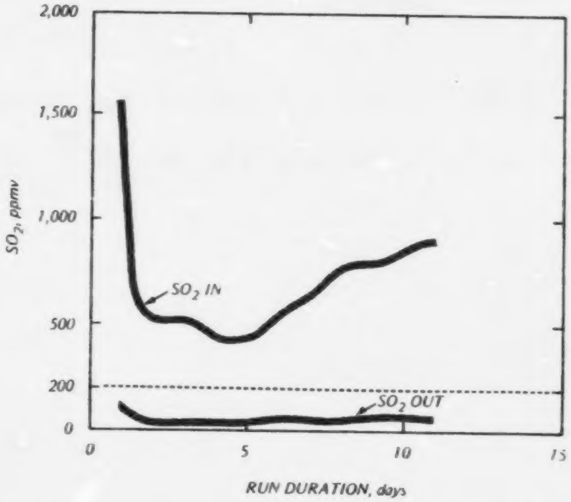
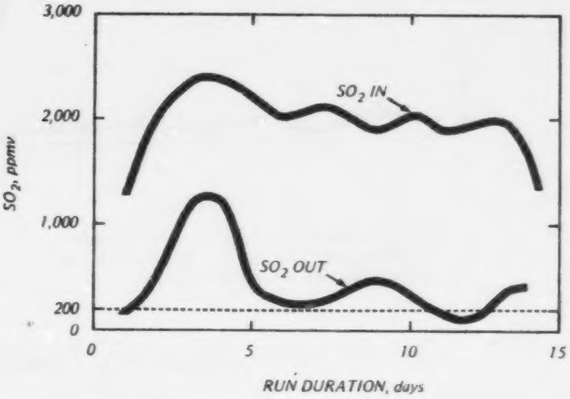


Figure 3-68. Inlet and outlet SO_2 concentrations during run no. 2.[81]



THE POOR SO_2 RECOVERIES DURING THIS PERIOD RESULTED FROM POOR QUALITY SOLUTION CAUSED BY MECHANICAL PROBLEMS IN THE SODA ASH FEED SYSTEM AND EVAPORATION AREA, AND LOW FEED RATES TO THE ABSORBER WHILE BALANCING TANK INVENTORIES.

Figure 3-69. Inlet and outlet SO_2 concentrations during run no. 3.[82]

by erroneous signals from a NIPSCO-installed pressure switch. Several minor boiler-related and FGD system problems occurred, but they caused no SO_2-removal difficulties.

The only other interruption of the integrated (Wellman-Lord and Allied units) operation was an 8-hour outage of the Allied sulfur-recovery unit, caused by excessive pressure drop across the reactor. This caused no problem in SO_2 removal, however, because of the storage capacity of the absorber surge tanks.

The inlet flue gas contained as high as 2800 ppm SO_2; the normal inlet SO_2 concentration, however, ranged from 2100 to 2300 ppm. Flue gas volume exceeded that expected. Outlet SO_2 concentrations normally ranged from 170 to 190 ppm, demonstrating the capability of SO_2 removal in excess of 90 percent. During one 4-hour period, SO_2 removal efficiency was 89 percent; and over the same period at the end of the acceptance trial a removal efficiency in excess of 90 percent was obtained.

For both the 12-day run at 544,000 actual m^3/hr (320,000 acfm) (92 MW equivalent) and the 83-hour run at 662,000 actual m^3/hr (390,000 acfm) (110 MW equivalent) the average SO_2 removal efficiency was 91 percent. The average SO_2 emission was 0.002 kg/1000 million J (0.5 lb/million Btu input), which is 42 percent of the current new source performance standard (NSPS) of 0.0055 kg SO_2/100 million J input) (1.2 lb SO_2/million Btu input).

The makeup chemical requirement guarantee was 5988 kg (6.6 ton) of soda ash (Na_2CO_3) per day. The actual usage was 5625 kg (6.2 ton) of soda ash per day, 94 percent of the guaranteed maximum.

The absorber was designed for a flue gas stream containing 5.2 percent excess oxygen, resulting in a purge of sodium sulfate of 6350 kg (7 ton)/day. Actual excess oxygen during the acceptance trial was 7 to 7.5 percent, with a maximum of 8.5 percent. The average sodium sulfate purge was 4082 kg (4.5 ton) per day, 64 percent of that expected at design conditions although the actual excess oxygen was about 40 percent in excess of design.

Utilities consumption was guaranteed not to exceed $56 per hour (1972 base costs). Actual utilities consumption, when multiplied by the cost factors in the contract, was $43 per hour (1972 base costs), 77 percent of the guaranteed maximum.

The required 99.5-percent minimum purity for the recovered sulfur was met.

Operation problems/solutions - During the early operating period (July to August 1976), the FGD plant experienced two major problems: 1) insufficient temperature control of the low-pressure steam to the evaporator and reduction area, and 2) insufficient turndown capabilities in the absorber operation. A shutdown was scheduled to relocate the steam desuperheater and to modify the absorber valve trays. During the shutdown, an inspection of the absorber internals revealed corrosion on the bottom surface of the lower-stage collector tray, which had been installed without rubber lining, and a seal failure where the trays connect to the absorber walls. The shutdown was then extended so the corroded tray surface could be lined with rubber [4 mm (5/32 in.) of tip-top rubber] and the tray-to-wall sealant could be replaced.[84,85]

On two occasions during the operations period from July through November 1976 the absorber was required to remove particulate matter under the most severe conditions.

- During a 3-day period, NIPSCO experienced problems with the ESP on Unit 11. The ESP was completely out of service for about three hours, and operated with only partial removal for the rest of the 3-day period. The absorber continued to operate through the period, but at reduced solution flow rates in order to control the fly ash buildup in the system. Following a brief power plant shutdown to repair the ESP, the SO_2 removal unit was started and leveled out, reaching normal operating conditions in only 1 day.[86]

- On another occasion the flue gas from Unit 11 contained a high loading of unburned coal. The Wellman-Lord SO_2 Recovery Process continued to operate, removing the particulate coal, but at reduced SO_2-removal capabilities because of the lower absorber solution feed rate.[87]

The sodium sulfate concentration level in the feed solutions to the absorber was to be maintained at 7 to 8 percent. Because of mechanical problems in the purge treatment area, however, the sulfate level reached 13 percent. Problems were experienced with the purge dryer, which restrained the crystallizer operation and prevented the discharge of the dried sodium sulfate. The drive train, feed box, and distribution problems in the dryer were corrected. During the subsequent operating period an icy crystal buildup formed on the crystallizer walls, restricting sulfate crystal formation. Several piping and control changes were made to the refrigeration system and to the crystallizers, which lengthened the operating period and shortened the time needed to descale the crystallizers increasing the overall purge treatment operating cycle.[88,89]

A review of the operating problems from July through November 1976 and the corrective actions taken is given in Table 3-48.[90]

During the acceptance trial, the only major problem was

Table 3-48. PROBLEMS ENCOUNTERED AT THE NIPSCO WELLMAN-LORD INSTALLATION[93]

(July to November, 1976)

Problem	Solution
No absorber turndown. The absorber was to operate between 46 MW and 110 MW without dumping the liquid from valve trays.	The valve trays were leveled to within 3.2 mm (1/8 in.) across a distance of 7.62 m (25 ft). Some valve cap weights were changed and some valves were replaced with a different valve type.
The collector tray seals leaked at the walls.	The original seal material between the metal tray and the tile wall of the absorber failed. This material was removed and replaced with packing and silicone caulking.
Absorber roof and stack joint leaks occurred.	The gaskets between the top of the absorber wall and cover and between the reheat venturi and the stack were reinforced.
Temperature control of the low-pressure steam was inadequate.	The desuperheater on the low pressure steam line was relocated to give better steam saturation and temperature controls.
There was corrosion on the bottom surface of the lower collector tray.	The bottom surface of the collector tray (exposed to flue gas) was sand-blasted and lined with cured rubber.
SO_2 analyzer sample probes became plugged.	Newly designed, traced, and air purged sample probes have been installed, eliminating the plugging.
Piping changes were made to the stack reheat system.	The size of the ring header supplying natural gas to the four burners was increased, and new regulators were installed to maintain steady gas pressure to all four burners.
Low pressure occurred in emergency steam supply piping to the FGD plant.	NIPSCO removed the flow meter orifice in the emergency steam line.

the shut downs of the blower booster turbine because of an erroneous signal from a pressure switch. A NIPSCO-installed pressure switch, designed to protect the boiler and the ductwork from excessive positive or negative pressure, became plugged. When cooled rapidly by rain or gusting wind, the gas inside the tubing contracted and gave a false indication of negative pressure. This signal shut down the blower booster turbine and opened the bypass damper to the stack. The pressure switch was restored to proper operation and the problem was eliminated.

3.5.3 Description of Foreign Installations

Seventeen Wellman-Lord installations are operating in Japan (Table 3-49). As of mid-1976 all reported to be achieving SO_2 removal efficiency greater than 90 percent (some in excess of 98 percent) while maintaining an on-stream time greater than 98 percent for the absorption area. None of these units operates on coal-fired boilers.[91,92]

3.5.3.1 Japan Synthetic Rubber - The first Wellman-Lord unit on an industrial boiler was installed on two oil-fired (4.0 to 4.2 percent-sulfur fuel oil) steam boilers at the Japan Synthetic Rubber plant at Chiba. Each boiler generates 84 metric tons/hr (185,000 lb/hr) of steam. The inlet SO_2 concentration is 1500 to 2000 ppm; the outlet SO_2 concentration is less than 200 ppm. The flue gas rate is 211,000 m^3/hr (124,000 scfm).[94,95,96] A flowsheet of the unit is given in Figure 3-70.

The flue gas is washed by the prescrubber installed in the lower part of the three-stage absorption tower.[97]

The greater part of solids and SO_3 formed in the boiler is removed with the wash liquor. The wash liquor is filtered, blended with the Na_2SO_4 purge stream (discussed later), and neutralized, before discharge from the system.

Table 3-49. OPERATIONAL WELLMAN-LORD FGD SYSTEMS IN JAPAN

Company/location	Completion date	Feed gas origin	Gas flow, 100 m^3/hr	Gas flow, (scfm)	SO_2 concentration, ppm	Disposition of SO_2
Japan Synthetic Rubber, Chiba	August 1971	Oil-fired boilers	200	(124,000)	in 2,100 out 200	Sulfuric acid
Toa Nenryo, Kawasaki	August 1971	Claus plants	67	(41,000)	in 6,500 out 200	Recycle to Claus plants
Chubu Electric Power, Nagoya	May 1973	Oil-fired boiler (220 MW)	620	(390,000)	in 1,600 out 150	Sulfuric acid
Sumitomo Chemical, Sodeguara	November 1973	Oil-fired boiler	360	(225,000)	in 1,550 out 250	Sulfuric acid
Japan Synthetic Rubber, Yokkaichi	December 1973	Steam boiler (2% S fuel oil)	450	(280,000)	in 1,000 out 100	Sulfuric acid
Kashima Oil, Kashima	February 1974	Claus plants	30	(20,200)	in 11,000 out 200	Recycle to Claus plants
Toa Nenryo, Hatsushima	October 1974	Claus plant	17	(10,000)	in 18,580 out 250	Recycle to Claus plant
Toyo Rayon, Nagoya	December 1974	Oil-fired boiler	330	(218,000)	in 1,500 out 150	Sulfuric acid
Japan National Railway, Kawasaki	September 1974	Steam boiler (200 MW-3% S fuel oil)	700	(435,000)	in 1,500 out 45	Sulfuric acid
Mitsubishi Chemical, Mizushima	April 1975	Oil-fired boiler	628	(373,000)	in 1,500 out 150	Sulfuric acid
Kurashiki Rayon, Okayama	July 1975	Oil-fired boiler	410	(248,000)	in 1,500 out 150	Sulfuric acid
Fuji Film, Fujinomiya	1974	Oil-fired boiler	150	(89,000)	in 1,300 out 125	Liquid SO_2
Shin Daikyowa, Yokkaichi	December 1975	Oil-fired boiler	400	(253,000)	in 1,500 out 150	Sulfuric acid
Sumitomo Chemical, Niihama	February 1976	Oil-fired boiler	155	(91,000)	in 1,600 out 130	Sulfuric acid
Mitsubishi Chemical, Mizushima	August 1976	Oil-fired boiler	628	(390,000)	in 1,300 out 130	Sulfuric acid
Mitsubishi Chemical, Kurosaki	September 1976	Oil-fired boiler	530	(330,000)	in 1,500 out 75	Sulfuric acid
Tohoku Electric Power, Niigata	March 1977	Oil-fired boiler	380	(236,000)	in 1,000 out 100	Sulfuric acid

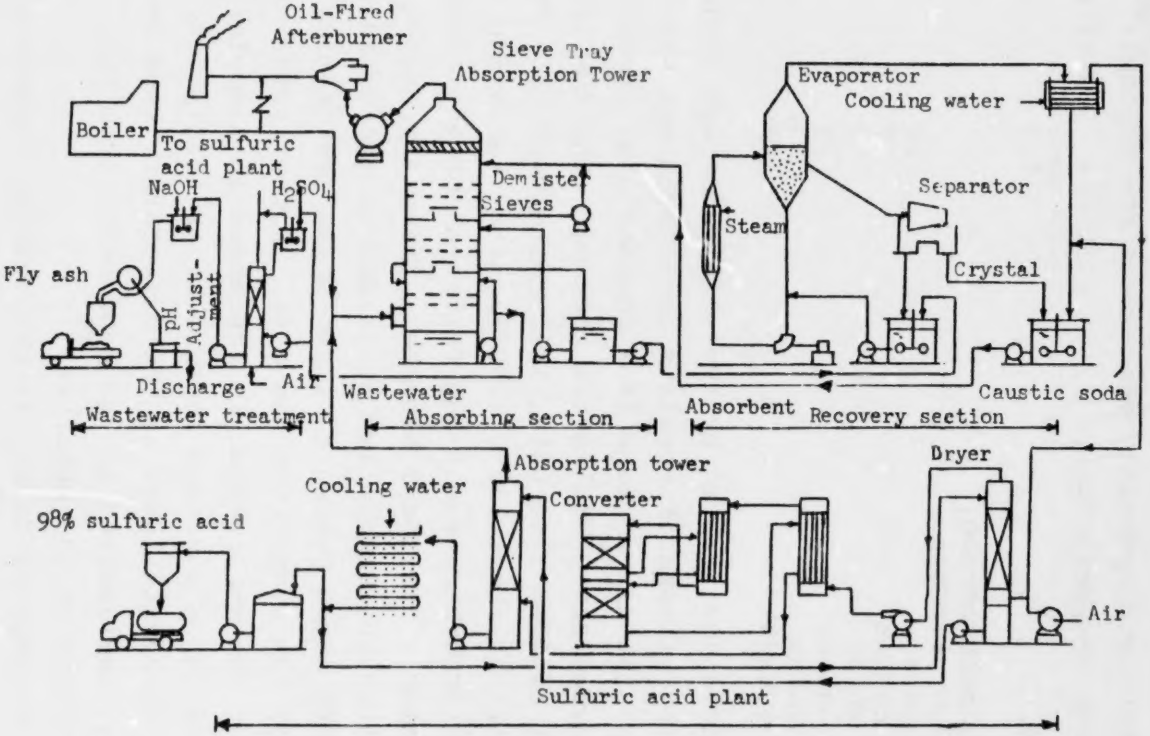


Figure 3-70. Flowsheet of Wellman-Lord process at Japan Rubber, Chiba, Japan.[99]

The partially cleaned gas rises in the tower while contacting a countercurrent flow of concentrated sodium sulfite solution that removes more than 90 percent of the inlet SO_2. The following scrubber operating parameters are typical:[98]

Gas velocity (superficial),	1.65 m/s (5.4 ft/s)
pH, inlet	7.3 to 7.4
outlet	5.5
Solution concentration, (percent by weight)	
Na_2SO_3, inlet	20
outlet	7
$NaHSO_3$, inlet	2
outlet	20
Na_2SO_4, inlet	5
outlet	>5

Mist droplets are removed by the eliminator and demister combination. The eliminator consists of a two-stage vertical section housing; the first stage is a conventional chevron-type and the second has a Tellerette packing. The demister is a horizontal section also using Tellerette packing. The gas is discharged at about 138°C (280°F) after reheating in an oil-fired afterburner, requiring 5 percent of the total energy consumption. The spent absorber, primarily $NaHSO_3$, is discharged from the absorption tower and is stored in a surge tank before being pumped to a two-stage evaporator.[100]

Gas leaving the evaporator has a SO_2 concentration of 90 percent after the water vapor is condensed in a cooler. The concentrated SO_2 gas stream is sent to an acid plant. Sodium sulfite is crystallized in the evaporator, separated by centrifuging, redissolved in the condensate from the cooler, and recycled to the absorber.[101]

To reduce buildup of Na_2SO_4 in the system, a portion of the mother liquor from the centrifuge is purged. This purge liquor is added to that from the cooler. The bleed taken off in the cooler is required to remove contaminants (sodium pyrosulfite, thiosulfate, dithionate, etc.) which would build up in the system. The purge streams undergo wastewater treatment. The final wastewater is an essentially clear, concentrated solution of sodium, which is discharged to the bay.[102]

This Wellman-Lord unit is noteworthy for its successful operating history. It started operation in June 1971 and operated at 97-percent availability during its first year and 100 percent during the second and third years. More than 90 percent of the SO_2 is removed, and only four hours of an operator's time are required per 8-hour shift.[103,104,105,106]

3.5.3.2 Chubu Electric Power - The first large installation of a Wellman-Lord unit at a Japanese utility plant came on line in late spring of 1973 at the Nagoya Station of the Chubu Electric Power Company in Japan. The plant has the capacity of treating 663,000 m^3/hr (390,000 scfm) of flue gas, containing 2100 ppm SO_2, from a 220-MW boiler which burns oil containing 3-percent sulfur. It was designed and constructed by Mitsubishi Kakoki Kaisha, a licensee of Davy Powergas. This is a peak-shaving power station. The FGD system was contractually required to handle stack gas fluctuations from 35 percent to 105 percent of design flow in 22 minutes, while maintaining outlet SO_2 emissions at less than 150 ppm. In May 1973 the outlet SO_2 was tested by the Japanese government at 130 to 135 ppm, which represents over 93-percent SO_2 removal.[107,108]

The process flowsheet and the plant layout are shown in Figures 3-71 and 3-72.

The flue gas from the boiler is cleaned in an ESP and precooled to 58°C (136°F) before entering the main part of the Wellman-Lord absorber. The L/G ratio is 0.7 l/nm^3 (5.2

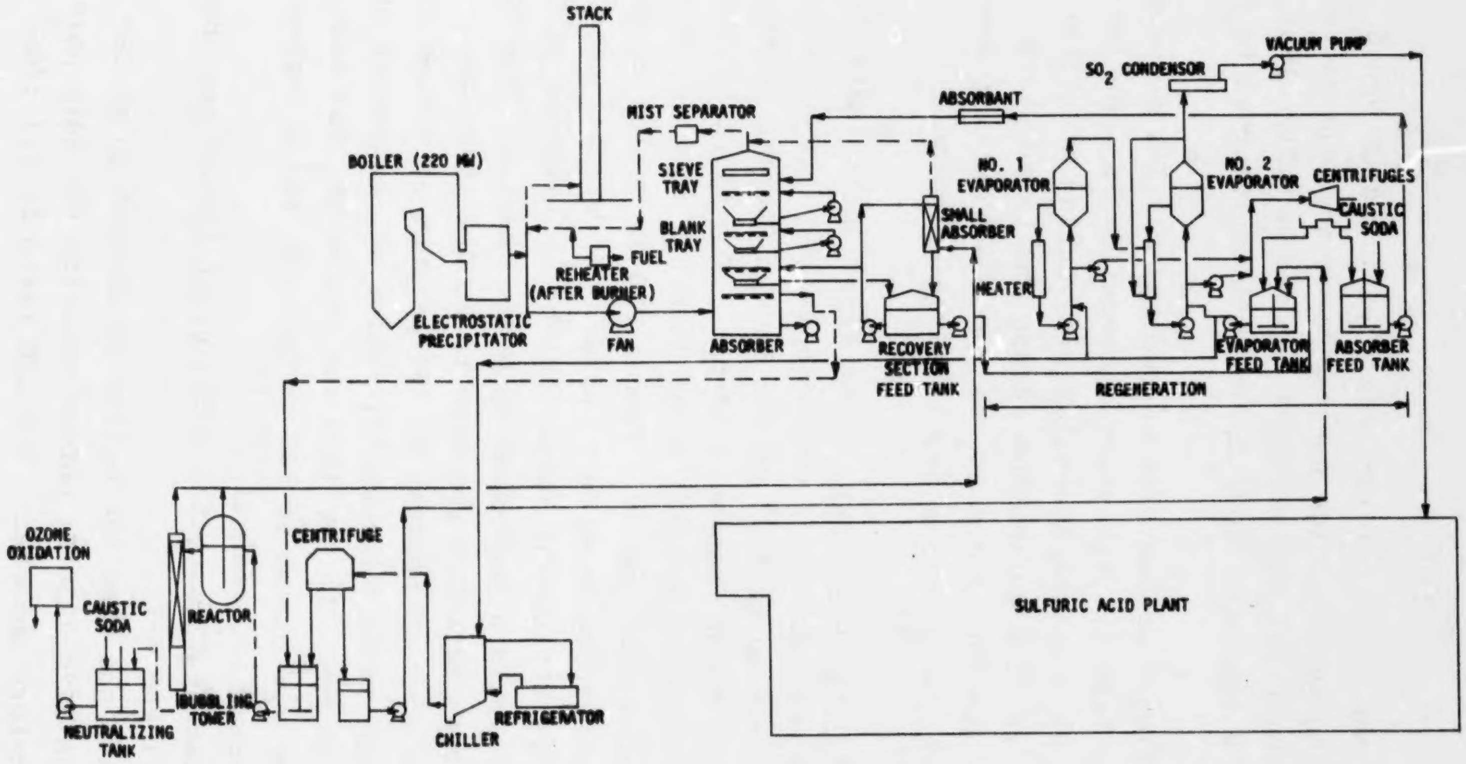


Figure 3-71. Flowsheet of the Wellman-Lord (MKK) process at Chubu Electric, Nagoya, Japan.[109]

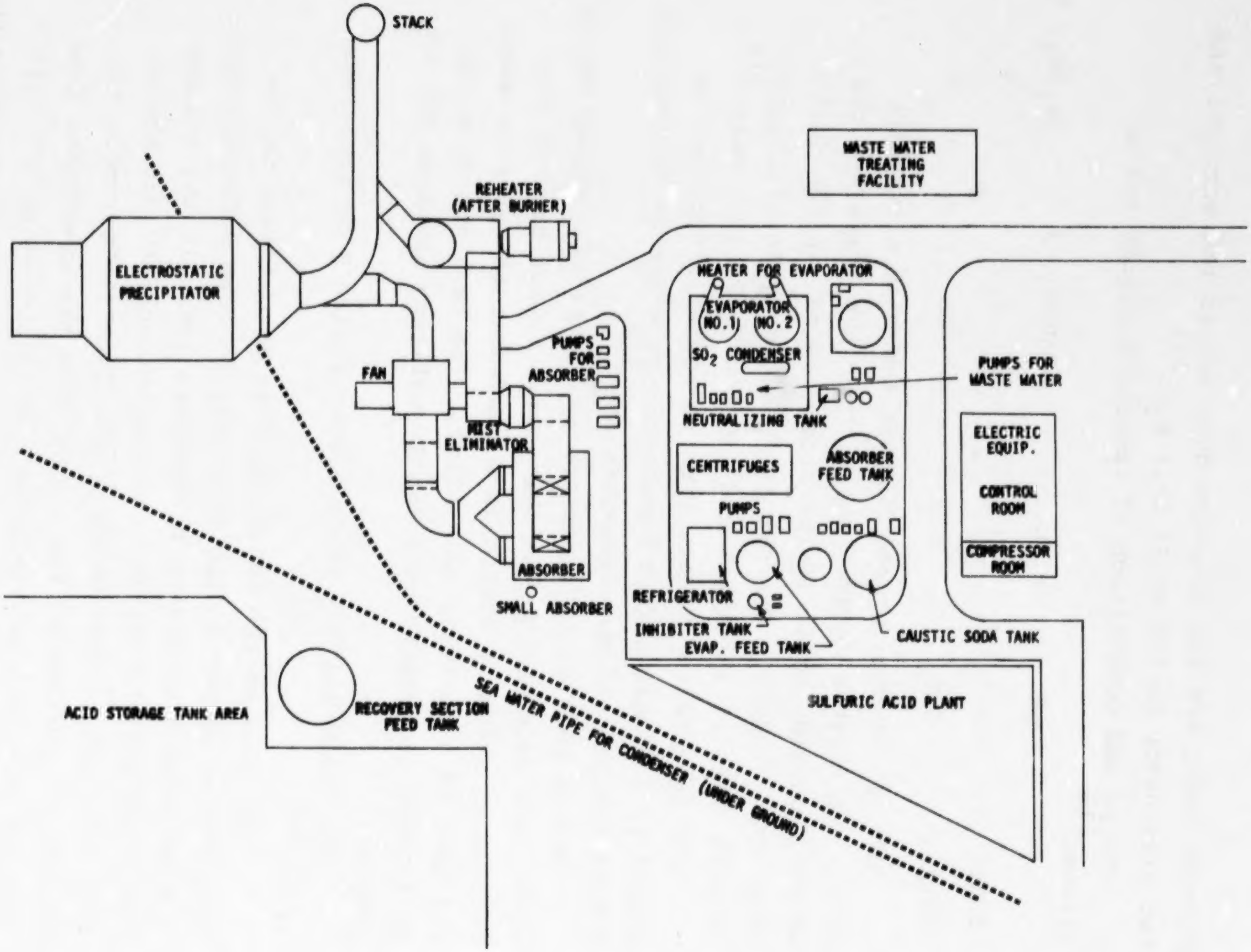


Figure 3-72. Layout of Wellman-Lord FGD plant at Chuba Electric, Nagoya, Japan.[110]

gal/1000 scf), and the pressure drop across the scrubber and mist eliminator is 500 mm (2 in.) H_2O.[111]

The pH and composition of absorbing liquor are as follows:[112]

	pH	Na_2SO_3, %	$NaHSO_3$, %	Na_2SO_4, %
Inlet	7.3-7.5	20	3	5
Outlet	5.5	7	22	5

The gas is then passed through two mist eliminator units, the first a chevron above the sieve trays and the second a packed vessel (Tellerette packing) in the dust after the scrubber. The gas is reheated by an afterburner to 140°C (284°F) in winter and 110°C (230°F) in summer.[113]

The scrubber effluent, rich in sodium bisulfite, is heated in a double-effect evaporator by steam to crystallize sodium sulfite and to regenerate SO_2.[114]

Sodium sulfite is separated by centrifuge, dissolved in water, and returned to the absorber. Sulfur dioxide gas containing steam is cooled in a condenser to separate water and sent to a sulfuric acid plant. The tail gas from the acid plant is treated by a small auxiliary scrubber for SO_2 removal.[115]

The Chuba unit is built almost entirely of stainless steel.[116]

A portion of the sodium sulfite is oxidized in the scrubber by oxygen in the flue gas, to form sodium sulfate. A small amount of sodium thiosulfate forms in the evaporator. Neither the sulfate nor the thiosulfate can absorb SO_2, and must be removed from the liquor. To remove the sulfate, a portion of the liquor from the evaporator feed tank is sent to a crystallizer; it is cooled to 0°C (32°F) to crystallize the sulfate, which is then separated from the liquor by a centrifuge. Liquor from the centrifuge is returned to the feed tank. Sulfate crystals containing a small amount of sodium sulfite and bisulfite are dissolved in a purge stream and combined with a small amount of sodium sulfite and bisulfite, purge containing thiosulfate contaminant. The mixed liquor is treated with sulfuric acid to decompose the sulfite, bisulfite, and thiosulfate; it is then aerated (the SO_2 released is sent to the auxiliary scrubber), oxidized with ozone for further decomposition of the thiosulfate, settled to precipitate solids, neutralized with sodium hydroxide, and then discharged into the sea along with other wastewater from power plants. The chemical oxygen demand (COD) is kept below 10 ppm, as required by regulations.[117]

When the unit was originally put on line, the operators inadvertently switched from low-sulfur (0.7%) to high-sulfur (4%) fuel oil; however, there was little effect on the outlet SO_2 concentration, and because of the automated adjustments to the process equipment at no time did it exceed 150 ppm.[118]

Ample absorber liquor surge capacity allows the evaporator and acid plant to continue operation at minimum load at weekends when the boiler and main scrubber are shutdown.[119]

The SO_2 gas removed in the system is cooled in a condenser to remove water, and then used for acid production in a sulfuric acid plant with a capacity of 73 metric ton/day (80 TPD) was built as part of this installation.[120]

The plant had some operating problems at first, including vibration of the sieve assembly, difficulty in controlling the draft at the I.D. fan at low load, and high turbidity of the by-product acid all these, however, were solved in a few months.[121]

The Nagoya plant has proved reliable and easy to control. The system is highly automated; only two operators are required for the entire operation, including start-up and shutdown. Automation is necessary, owing to the widely fluctuating load and the fact that the boiler is shut down every weekend.[122,123]

3.5.4 Engineering Design Parameters

Sulfur dioxide removal efficiencies in excess of 90 percent have been achieved with all the operating Wellman-Lord FGD systems. Removal efficiencies in excess of 98 percent have been reported at Japanese units; they have exceeded 97 percent at sulfuric acid and Clauss operations in the U.S.; and 90 percent at the acceptance trials completed on September 14, 1977, at NIPSCO.[124,125,126,127,128]

To reduce the pressure drop required across the prescrubber, and to maintain good prescrubber mechanical reliability, Davy Powergas recommends the use of an ESP for primary particulate control upstream of the prescrubber and absorber. The prescrubber provides additional particulate removal and saturation of the flue gas to the adiabatic temperature (approximately 49°C to 52°C, 120°F to 130°F).[129,130]

The prescrubber may be constructed of rubber-lined mild steel with acid-proof brick lining. In Japan, Nippon Steel 45M stainless steel (equivalent to 329 SS) was used for the prescrubber. The L/G in the prescrubber has varied as a function of particle size and the pressure drop. The L/G at NIPSCO ranges from 800 to 1070 $l/1000\ m^3$ (6 to 8 gal/1000 acf) at 1 kPa (4 in. H_2O) pressure drop.[131,132]

According to Davy Powergas, the L/G ratio for each tray of the absorber ranges from 200 to 535 $l/1000\ m^3$ (1.5 to 4.0 gal/1000 acfm). Davy Powergas currently favors about 260 $l/1000\ m^3$ (2 gal/1000 acfm). Each absorber will have three (for high inlet SO_2 concentration) to five (low inlet SO_2 concentration) valve and collector trays. The overall L/G for the absorber in a new installation will range from 800 to 1350 l/1000 m^3 (6 to 10 gal/1000 acfm).[133]

The pH of the sodium sulfite absorbing solution is maintained at approximately 6.0 at the absorber solution inlet (in new designs, in the top tray of the absorber). The pH of the purge stream from the absorber ranges from 4.4 to 4.5.[134]

The system makeup feed of fresh, 20-percent sodium carbonate solution is approximately 67 l/1000 m^3 of flue gas (0.5 gallon/1000 acfm) per tray. All the fresh solution is fed onto the top tray, so that chemical equilibrium is not established and no theoretical lower limit to the exit SO_2 level exists.[135]

For the valve tray assemblies within the absorber, the design superficial gas velocity is 2.7 to 3.1 m/s (9 to 10 ft/s). The pressure drop across each absorber tray is approximately 0.87 kPa (3.5 in. H_2O). For three to five trays, the pressure drop across the absorber itself is 2.6 to 4.4 kPa (10.5 to 17.5 H_2O).[136,137]

The absorber trays are Koch Multiventuri Flexitrays (MYFT). The valve trays may be constructed of 316 SS, 316 L SS, or Hastalloy G. Normally, Hastalloy G would only be used for the bottom tray, where the most acidic conditions occur.[138,139]

Scaling in the SO_2 absorber has never been reported as a problem in any Wellman-Lord installation.[140,141]

The current maximum-size booster fan/scrubber/absorber module which Davy Powergas recommends will handle approximately 8500 m^3/min (300,000 scfm) of flue gas.[142,143]

As the SO_2 inlet concentration increases, the number of mass transfer stages required to maintain a desired efficiency may decrease. This is due to the higher driving force of the more concentrated SO_2 gas stream. For example, for 90-percent removal, an inlet SO_2 concentration of 1000 ppm may require as many as five trays or mass transfer stages (e.g., San Juan). For the same 90-percent removal, an inlet SO_2 concentration of 3000 ppm may require only three mass transfer stages (e.g., D.H. Mitchell). Each mass transfer stage will add 0.6m to 1.5m (2 to 5 ft) to the absorber height. Each additional stage added to the absorber accounts for an approximate 10-percent increase in capital cost.[144,145,146]

The water content of the flue gas for a given dry bulb temperature will have a direct effect on the number of stages required to satisfy the SO_2 removal requirements. The absorber will operate 2°C to 7°C (4°F to 12°F) higher than the adiabatic saturation temperature of the flue gas. Since this operating temperature of the absorbing section will affect SO_2 equilibrium, a 2-volume-percent change in water content of the flue gas may require one additional mass transfer unit, since the saturation temperature will be higher.[147]

The chloride level in the absorber is maintained in the range from 200 to 300 ppm to avoid corrosion of the stainless steel internals. The chloride level is controlled by the purge rate from the absorber and by the prescrubber.[148]

The key component of the sulfur recovery area is the forced-circulatory, double-effect evaporator. This is constructed primarily of 316 L SS. The majority of the remaining chemical plant components (partial evaporators, condensate stripper, storage tanks, and related pumps and piping) are constructed primarily of 316 and 316 L SS and, where possible, of mild steel.[149]

The chloride level in the mother liquor stream may be in the 2000-ppm range. This level has been measured at NIPSCO without any observed serious damage to the components. Again, the chloride level is controlled by removal in the sulfate purge stream.[150]

Several side-products arise from thermal regeneration of sodium bisulfite solution in the evaporator to sodium sulfite. For simplified reporting these products are often expressed as thiosulfates. The thiosulfate level is normally maintained from 0.2 to 0.3 percent by careful temperature and pressure control in the chemical plant and in the bleed/purge stream. If the thiosulfate were permitted to build up beyond the control range, it could break down to elemental sulfur. Over an extended period, the sulfur buildup could foul the absorber and evaporator surfaces.[151]

The percentage of solids in the evaporator/crystallizer is a control variable for the sodium sulfate regeneration process.

The evaporator is often designed for 45 to 50 percent solids (sulfite crystals). Evaporators have been operated for prolonged periods at approximately 60 percent. The limiting factors are formation of crystals on the heat transfer surfaces and the viscosity, which limits the ability to recirculate the slurry.[152]

The evaporator may be either a single-or a double-effect type. If less than 2268 kg (5000 lb) of SO_2 per hour is being handled, a single effect evaporator is normally chosen. The operating conditions for a single effect evaporator are 55 to 69 kPa (8 to 10 psia) operating pressure and 88° to 99°C (190° to 210°F). A double-effect evaporator is normally chosen above 2268 kg (5000 lb) of SO_2/hr. The operating conditions are 27 to 41 kPa (4 to 6 psia) operating pressure and 71° to 79° (160° to 175°F).[153,154]

The storage or surge tanks in the chemical regeneration plant are often designed with the storage capacity for 8 to 24 hours of full process stream flow. This enables the regeneration plant to be shut down for maintenance without interrupting SO_2 removal in the absorber section. The necessity for extensive installed spare equipment is therefore reduced without sacrificing control reliability.[155,156,157]

The sulfate purge from the Wellman-Lord system can amount to 5 to 10 percent of total sulfur removal.[159,160,161] The size of the purge stream has increased with the increasing size of new Wellman-Lord installations. Attempts to prevent solution oxidation by the addition of antioxidants were made, but were abandoned because of high cost and the fact that purge would still be required. Selective removal of the oxidized portion of the solution (sodium sulfate) by chilled-wall crystallizers has been more successful, resulting in a five-to six-fold decrease in the purge stream and the chemical makeup requirements.[162,163]

As higher SO_2 removal efficiencies are required, more space is required for the chemical plant. One good aspect of the Wellman-Lord process is that the absorption and chemical plant areas can be separated as far apart as 800 m. (0.5 mile) without any major capital or operating cost increase.[164]

3.5.5 Process Operability

In the United States, the six operating Wellman-Lord units installed on sulfuric acid or Claus sulfur recovery units are reported to have an on-stream time of greater than 97 percent for the absorption area. In Japan in mid-1976 the approximately thirteen Wellman-Lord units on oil-fired boilers and Claus sulfur recovery units maintained an on-stream time greater than 98 percent for the absorption area. During this time SO_2 removal efficiencies of these units were reported to exceed 90 percent, with some units reportedly experiencing 97-to 98-percent SO_2 removal.[165,166,167,168,169]

The Wellman-Lord process has had SO_2 outlet concentrations guaranteed in the range from 200 ppm (NIPSCO) to 70 ppm (San Juan 1 and 2).[170,171] The corresponding removal efficiencies would be 91 and 90 percent respectively. Design outlet SO_2 concentrations within this range have been met and have been below design in a number of cases. Most recently, the SO_2 removal guarantee, the chemical usage guarantee, and the utility usage guarantee have been met during the acceptance trial at NIPSCO (8/29/77 through 9/14/77).[172]

Davy Powergas, the process vendor, when acting as primary contractor, is willing to guarantee the minimum percentage SO_2 removal (e.g., 90-percent SO_2 removal) and/or the maximum SO_2 outlet concentration (e.g., 200 ppm at NIPSCO), the chemical usage, the utility usage, the particulate removal, and other items that the individual client may request, depending upon the project parameters.[173]

The only corrosion problem has been in the prescrubber/absorber area. Use of rubber-coated 316 L SS or Hastalloy G (44 percent Ni, 22 percent Cr, 20 percent Fe, 6.5 percent Mo, 2.1 percent Cb + Ta, 2 percent Cu, and 0.05 percent C max.) for the lowermost valve and collector tray protects against the acidic conditions encountered there. At NIPSCO, the rubber coating of the collector tray caused no unscheduled shutdown.[174,175,176]

No scaling has been reported in any of the 24 operational Wellman-Lord absorber units in the United States or Japan.[177,178]

The ability of several Wellman-Lord systems to adjust to inlet flue gas composition variations is as follows:

- At the automated Wellman-Lord unit at Chubu Electric Power, Nagoya, Japan, shortly after the unit went on line the operators inadvertently switched from low-sulfur (0.7 percent) to high-sulfur (4 percent) fuel oil. The outlet SO_2 concentration had been 130 to 135 ppm; after the changeover, automated adjustments to the process equipment kept the SO_2 emission below 150 ppm.[179]
- During a 3-day period, NIPSCO experienced problems with the ESP on Unit No. 11. The ESP was completely out of service for about 3 hours, and operated with only partial removal for the rest of the 3-day period. The absorber continued to operate through the period, but at reduced solution flow rates to control the fly ash buildup in the system. Following a brief power plant shutdown to repair the ESP, the SO_2 removal unit was restarted and leveled out, reaching normal operating conditions in only 1 day.[180]
- On another occasion, the flue gas from Unit No. 11 contained a high loading of unburned coal. The Wellman-Lord SO_2 Recovery Process continued to operate, removing the particulate coal, but penalizing the SO_2 removal capabilities because of the lower absorber solution feed rate.[181]

The Wellman-Lord units may produce sulfur, liquid sulfur dioxide, and/or sulfuric acid as salable products. The market for sulfur and/or sulfuric acid must be considered in the decision as to the form in which the SO_2 will be recovered. Often the product of choice is sulfuric acid.

The by-product of the Wellman-Lord process is sodium sulfate, which is often dried for sale or for disposal. The possible market for sodium sulfate or possible disposal problems must be investigated with regard to the location of the individual Wellman-Lord unit.[182,183,184,185]

3.5.6 Pilot Plant and Prototype Potential

All inactive sodium salts must be purged from the scrubber solution and their sodium ions replaced by active sodium ions from soda ash or caustic soda. Research is being conducted into methods for eliminating the purge stream by reactivating the sodium ion. A number of processes have been investigated in both laboratory and full-scale applications, but few details have been disclosed. Davy Powergas hopes to close the loop, by reactivating as much of the inactive sodium ion as is economically feasible and by generating sodium carbonate. In one process, the purge stream would be incinerated to liberate the SO_2 and recover the sodium salts, a concept similar to the recovery boiler used in the pulp and paper industry. Two processes are under development to reduce the sodium sulfate chemically, one of which uses natural gas and the other coal as the reductant. Pilot plant testing has been successful. A method employing high-temperature separation of the sulfate has also been successfully tested on bench scale.[186,187,188]

The shortage of natural gas could be a major limitation on the use of the Wellman-Lord process to recover elemental sulfur. A 500-MW plant would require about 23 m^3/min (800 scfm) of natural gas, depending upon the sulfur content of the coal and the efficiency of the recovery plant. High-sulfur residual oils, coal gasification products, and alternative hydrocarbon fuels such as propane have been found to be useful as reducing compounds.

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117. Ibid., pp. 5-7 and 5-8.

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151. Ibid.

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183. Op. cit. No. 4, p. 462.

184. Op. cit. No. 13, p. 759.

185. Op. cit. No. 53. pp. 159-160.

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3.6 OTHER FGD SYSTEMS

This section describes other FGD systems that are not widely used in the United States.

3.6.1 Chiyoda Thoroughbred 101 Process[1,2,3,4]

In the Chiyoda process, SO_2 is absorbed in weak sulfuric acid and then oxidized by air to sulfuric acid. Part of the acid is neutralized with limestone to produce gypsum; the remainder is circulated as SO_2 absorbent. Thirteen utility oil-fired boilers in Japan, up to 350 MW in capacity, use the process with greater than 90 percent SO_2 removal.[5,6] A 23-MW prototype unit has been tested on a coal-fired boiler at the Scholz Station of Gulf Power Company. No full-scale systems have yet been installed on any U.S. plants.

The Chiyoda process is a variation of the double alkali process, which uses a clear solution in the scrubber. The process has four basic operations: prescrubbing, absorbing, cyrstallizing, and by-product/waste handling (Figure 3-73). Following removal of most of the particulate matter by means of an ESP, flue gas is quenched and cooled in a venturi-type prescrubber, where fly ash and chlorides are removed. Solids in the prescrubber liquid are removed in a thickener and the liquid is recycled. Particle-free gas proceeds to a fixed-bed scrubber, where SO_2 is absorbed in a countercurrent flow of weak sulfuric acid solution, containing a ferric ion catalyst, to produce sulfurous acid. Flue gas from the scrubber passes through a mist eliminator and reheater, and is then discharged. Continuous withdrawal of the absorbent maintains scrubber acid content at about 2.0 percent by weight. In a subsequent oxidation tower, air is bubbled through the scrubber effluent, catalytically oxidizing the SO_2 (by the ferric ion) to sulfuric acid. Part of this liquor is neutralized with lime or limestone to produce gypsum; the rest is recycled to the absorber. Chiyoda has developed a special gypsum crystallizer that produces large crystals. Smaller gypsum crystals from the overflow at the crystallizer and of the centrifuge overflow are settled out in a clarifier and used as seed crystals. Clarifier overflow is returned as mother liquor to the absorber. A portion of this stream is purged to reduce the buildup of soluble matter, such as chlorides. The basic chemistry of the Chiyoda process is as follows:

$$H_2O + SO_2 \rightarrow H_2SO_3 \quad (1)$$

$$H_2SO_3 + 1/2O_2 \xrightarrow{Fe++} H_2SO_4 \quad (2)$$

$$H_2SO_4 + CaCO_3 \rightarrow CaSO_4 + H_2O + CO_2 \quad (3)$$

Based on Japanese operating experience with oil-fired boilers, the Chiyoda 101 process has demonstrated a number of advantages: relative simplicity, a high degree of reliability, and the production of a high-quality gypsum by-product. In addition, scaling is prevented in the absorber/oxidizer or mist eliminator, since the scrubber liquid is not saturated with calcium sulfate. Sulfur dioxide concentration in the tail gas can be 50 ppm or less, representing 95 percent SO_2 removal.[8] The main drawback of the process is the high cost associated with its high liquid/gas ratio. The weak acid solution cannot absorb and retain much SO_2 per unit volume, and an L/G of 40 l/m^3 (300 gal/1000 acf) for a 1000-ppm SO_2 stream is required for 90-percent removal.[9]

Chiyoda has developed a variation of this process for their simultaneous NO_x-SO_2 removal method known as Process 102. Ozone, added to the entering gas stream, partially oxidizes NO_x after prescrubbing and cooling. Both SO_2 and

NO_x are then absorbed in the FGD unit, with part of the NO_x being converted to nitric acid and the remainder being released, largely as nitrogen and nitrous oxide, with the flue gas. Sixty-percent NO_x removal has been reported in a Japanese pilot plant. No data on SO_2 removal capabilities of Process 102 were found.

The first test of the Chiyoda 101 process on a coal-fired boiler was conducted under the direction of Southern Services, Inc., on Gulf Power Company's 23-MW Scholz plant at Sneads, Florida. The system began operation in February 1975. Total availability for the remainder of 1975 was 60 percent.[10] The system broke down for 6 weeks in August and September, after which it operated at greater than 97-percent availability to the end of the year.[11] Typical SO_2 removal was 82 to 87 percent with relatively low-sulfur coal (1.7 or 2.2 percent).[12] The gas rate varied between 16.5 and 21.2 m^3/sec (35,000 and 45,000 scfm). Operation of the system was terminated in March 1977.

The major operating problems occurred when both centrifuges became unbalanced because of a high, nonuniform wear on the screw conveyors. The hard facing on the screws wore much faster than was suggested by the Japanese experience, possibly because a different facing technique was used. Replacement of the screws alleviated the problem. Although there were other, more common equipment problems, no scaling occurred in the absorber, oxidizer, or mist eliminator. Liquid waste streams from the plant were combined, neutralized with limestone, and discharged to a settling pond, whose overflow went into the plant ash pond and eventually to a river.

3.6.2 Citrate Process[13,14,15,16,17]

The citrate process for flue gas desulfurization is a buffered aqueous absorption process that uses sodium citrate,

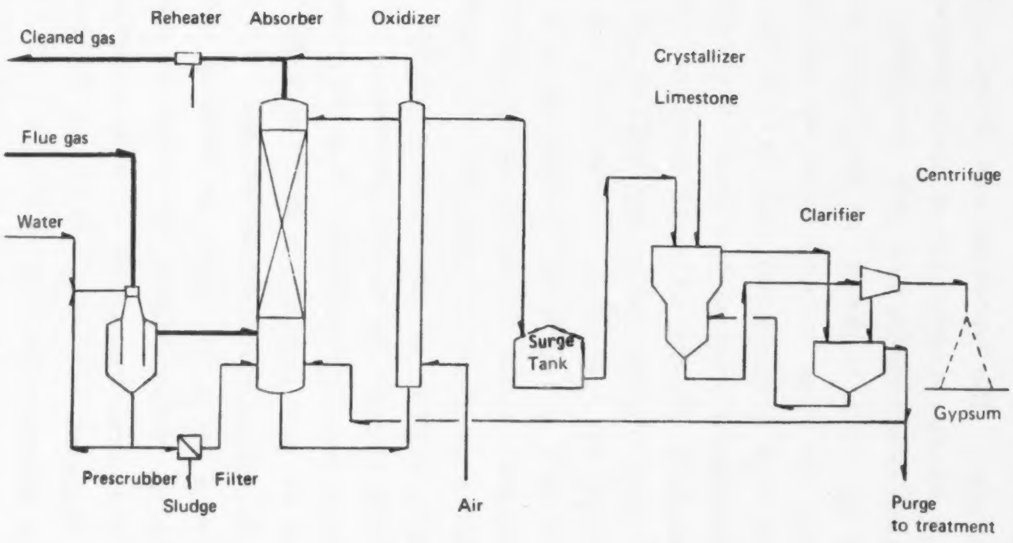

Figure 3-73. Flowsheet - Chiyoda Thoroughbred 101 process.[7]

or sometimes phosphate, as the buffering agent. The SO_2 is absorbed in solution, then reacted with gaseous H_2S to precipitate sulfur and regenerate the solution for recycling. Elemental sulfur is precipitated directly from the scrubber solution. The process was developed mainly by the U.S. Bureau of Mines; significant research was also conducted by Pfizer-Peabody-McKee and Stauffer. The different versions vary in such factors as absorber configuration, absorber liquor pumping rates, type of buffer, operating pH, and methods of H_2S production and sulfur separation. A partially integrated system has been demonstrated.

The chemistry of the citrate process is very complex and not yet fully understood. By buffering the absorption solution at a pH of 3.0 to 3.7 with various organic acids, principally citric acid, side reactions are controlled and sulfur is produced without building up concentrations of undesirable side products. Citrate suppresses side reactions, yielding sulfur as the main product. Figure 3-74 illustrates the Bureau of Mines process; it is representative of various configurations. The SO_2-bearing gas is first cleaned of particulate matter and sulfuric acid mist, then cooled to saturation temperature. It is absorbed in an absorption tower by a solution of sodium citrate, citric acid, and sodium thiosulfate, to produce sodium sulfite. Liquor from the absorber goes to a regeneration reactor, where it is reduced with H_2S, both to produce elemental sulfur and to regenerate the solution. Sulfates are purged from the liquor by cooling and crystallizing Glauber's salt ($Na_2SO_4 \cdot 10H_2O$), which is removed by filtration and then washed. The sulfur is separated from the solution by oil flotation and melting, and the regenerated solution is recycled to the absorber. The Bureau of Mines process uses

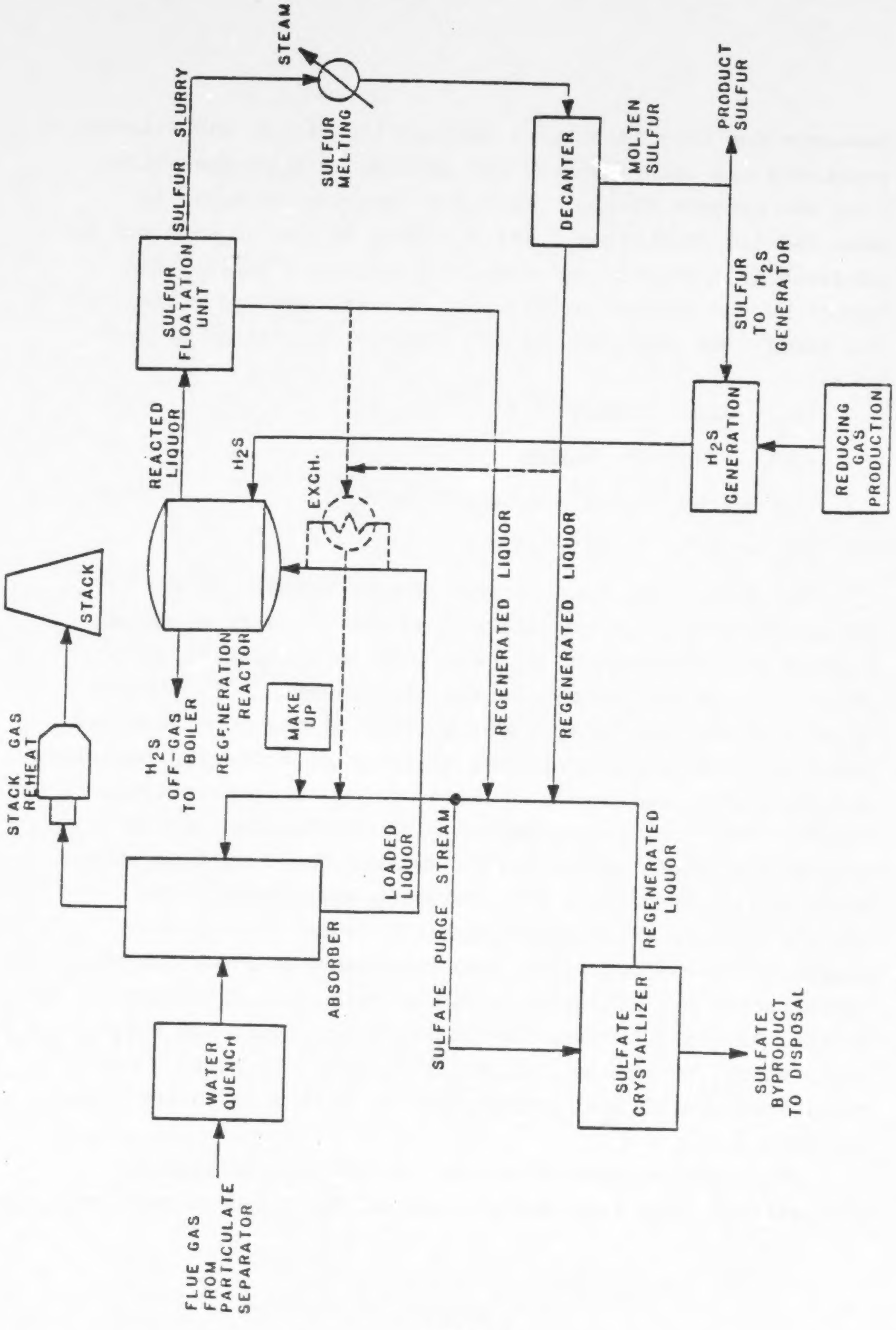

Figure 3-74. Flowsheet - Citrate process.[18]

kerosene for this flotation, whereas the Pfizer and Stauffer processes use other proprietary methods to separate sulfur from the product slurry. Flue gas from the absorber is reheated and discharged. The H_2S used in the process can be produced by a variety of methods, including reacting two-thirds of the product sulfur with natural gas and steam. The simplified chemistry of the citrate process is as follows:

$$SO_2 + H_2O \rightleftarrows HSO_3^- + H^+ \tag{1}$$

$$HSO_3^- + Na^+ \rightleftarrows NaHSO_3 \tag{2}$$

$$Cit^{=} \text{ or } PO_4^{=} + 3H^+ \rightleftarrows H_3Cit \text{ or } H_3PO_4 \tag{3}$$

$$SO_2 + 2H_2S \rightarrow 3S + 2H_2O \tag{4}$$

The major advantage of the citrate process is its direct precipitation of sulfur from the scrubber solution. Removal of 99 percent of SO_2 has been demonstrated, but there is no effect on NO_x in the gas stream.[19,20] Citrate is used as the buffering agent because of its good chemical stability, low vapor pressure, adequate pH buffering capacity, and the purity and physical character of the precipitated sulfur. The Stauffer process employs phosphate: it is cheaper and lasts longer than citrate. The resultant anhydrous Na_2SO_4 is said to be a valuable by-product. The citrate process is also efficient in terms of fuel consumption --the incremental fuel requirements for a 500-MW installation are estimated to be as follows: stack gas reheat, 23.1 x 10^6 J/sec (39 MM Btu/hour); heat for anhydrous Na_2SO_4 by-product, 11.4 x 10^6 J/sec (39 MM Btu/hour). Total incremental fuel consumption is 7.19 x 10^5 J/kWh (682 Btu/kWh).

There are certain drawbacks to the citrate process. Production of H_2S from methane and sulfur can be a problem, as a result of the shortage of natural gas (methane). Further, the relatively low pH in the scrubber can result in low efficiency, necessitating a large regeneration system, if high bisulfite formation is attempted. Since the absorber is also large, retrofitting existing boilers with citrate FGD systems may sometimes prove difficult. The separation of coal fly ash, which is necessary to avoid contamination of the product sulfur, further increases costs. There is also a potential problem with hydrocarbon losses from kerosene flotation.

Variations of the citrate process have been tested in three pilot plant applications: a Bureau of Mines, 0.5-MW plant on a slipstream from the Bunker Hill lead sintering furnace in Idaho; the Pfizer 1-MW plant on a coal-fired industrial boiler at Terre Haute, Indiana; and the Stauffer plant on a low-sulfur, oil-fired boiler at Northeast Utilities' Norwalk Harbor station in Connecticut, which used phosphate instead of citrate. All three plants showed good results. Kinetic models exist for scaling to larger sizes, but have not been demonstrated.

The Bureau of Mines citrate plant operated for 4500 hours between February 1974 and November 1975, producing more than 50 net tons of sulfur from a feed gas averaging 0.5 percent SO_2. The tail gas from the sinter plant was first cleaned in a baghouse, cooled in a packed wet scrubber, and then cleaned of acid mist and particulate in a wet ESP. To better simulate conditions in a coal-fired boiler, the gas steam was diluted to below 0.5 percent SO_2 (typically 0.07 to 0.13 percent). Kerosene or another hydrocarbon oil was added for the sulfur flotation-separation step; it was found that the use of air alone for flotation produced a wet sulfur product of about 8 percent solids. A glycolate scrubbing solution was satisfactorily substituted for citrate, at about half the cost, for about 450 hours of the pilot plant operation. Because of the irregular operation of the sinter plant, the longest continuous run was 265 hours. The sulfur product at the Bunker Hill pilot plant was of greater than 99.5 percent purity, with an ash content of about 0.03 percent. Carbon content was 0.2 to 0.3 percent with kerosene flotation, and less than 0.04 percent with air flotation.[21] With a gas flow of 0.47 m^3/sec (1000 scfm), SO_2 removal exceeded 99 percent and feed gas ranged from 70 to 550 ppm SO_2.[22] Removal efficiency dropped to 95 percent for a 130-ppm SO_2 gas at 0.61 m^3/sec (1300 scfm).[23]

A number of problems arose at the Bunker Hill plant. Incomplete regeneration of the absorbent solution was found to be caused by insufficient retention time of the H_2S gas in the single-surface precipitation reactor. This was solved by increasing contact time through the use of two reactors, addition of sodium thiosulfate to inhibit oxidation, and adding sodium carbonate to neutralize sulfuric acid in the solution. An early problem with cloudy recycle solution from kerosene flotation was caused by delayed precipitation of colloidal sulfur. It was corrected by increased contact time and by bleeding about 5 percent of the absorption solution to the reactor effluent tank, where it reacted with the excess absorbed H_2S. There were also some plugging problems, with sulfur in the pipelines between the reactors, for example.

A demonstration plant of the Bureau of Mines Citrate Process is currently under construction at a coal-fired power plant at the St. Joe Zinc Co. in western Pennsylvania.

3.6.3 Westvaco Activated Carbon Process[24,25,26]

The Westvaco process is a dry process featuring in situ oxidation of SO_2 while it is still in the gas stream.

Multistage fluidized beds are used for both adsorption of SO_2 and for regeneration of the carbon by hydrogen reducing-gas, simultaneously with the production of elemental sulfur. The major drawback is that it is a relatively energy-intensive process with high hydrogen consumption. Reaction heat is supplied by low-Btu gas from a gasifier or fuel oil. The fluidized-bed approach permits the handling of relatively large gas volumes. More than 90-percent removal has been demonstrated from a flue gas containing 2000 ppm SO_2.[27]

In the Westvaco process, the catalytic and sorptive character of carbon is exploited to convert SO_2 to sulfuric acid within carbon granules. Carbon promotes the necessary reactions, but does not directly take part in them. Figure 3-75 is a flow diagram of the process. The flue gas initially contacts dry granular activated carbon at stack gas temperature. Shallow-fluidized beds, the number of which are dependent on the boiler size, are used for 90-percent SO_2 removal. The SO_2 is removed from the gas through catalyzed oxidation to SO_3, followed by hydrolysis to sulfuric acid, which remains sorbed in the carbon granules. In a second fluidized-bed reactor, the carbon granules are reduced with hydrogen sulfide at 150°C (300°F), producing elemental sulfur and water vapor. Hydrogen sulfide is regenerated and elemental sulfur removed by a thermal stripping process near 540°C (1000°F) in a third fluidized-bed reactor with an external hydrogen source. The basic process chemistry is as follows:

$$SO_2 + 1/2O_2 + H_2O \rightarrow H_2SO_4 \qquad (1)$$

$$H_2SO_4 + 3H_2S \rightarrow 4S + 4H_2O \qquad (2)$$

$$3H_2 + 4S \rightarrow 3H_2S + S \qquad (3)$$

Because it is a dry process, the Westvaco process does not involve the critical control of chemistry necessary in

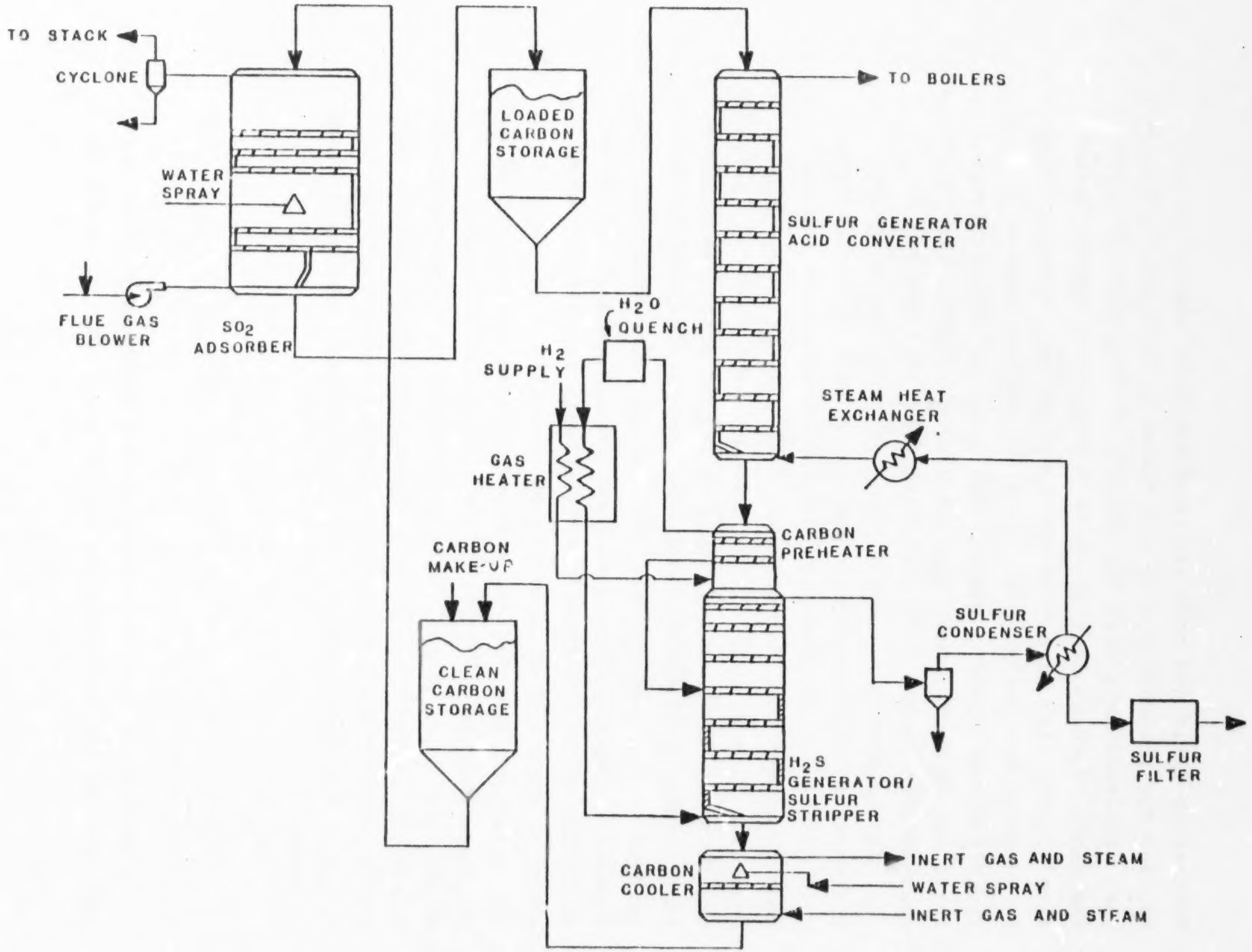

Figure 3-75. Flowsheet - Westvaco activated carbon process.[28]

wet reactions. Both the separation of water from by-products and the reheating of stack gas are eliminated. In addition, there is no sulfate problem and readily available coal may be gasified for use as the reductant. Nor is high-quality hydrogen essential; dilute gas from an air-blown gasifier can be used if the tars are removed and carbon monoxide is reduced to a low level. Removal of 99 percent of the SO_2 is possible with increases of only about 20 percent in adsorber size and pressure drop.[29] Estimated energy requirements for a 500-MW unit are: electric power, 13 MW; fuel oil, 79×10^9 J/hr (75 MM Btu/hr); steam, 68×10^8 J/hr (6.4 MM Btu/hr); and reducing gas, 18×10^{10} J/hr (175 MM Btu/hr). Total incremental fuel requirements are 10.6×10^5 J/kWh (1000 Btu/kWh) equivalent to an energy requirement of 50 MW.

Several potential design and operating problems are involved in scaling-up the Westvaco process to commercial size. Compared with most wet scrubbers, a fluidized-bed unit operates at a higher pressure and at a lower gas velocity: which means it must have a larger diameter and a higher power consumption. A 250-MW unit would require 2,323 square meters (25,000 ft^2), a size large enough to create problems for many retrofit installations. A hard, reactive carbon absorbent is necessary to minimize raw material costs and particulate emissions. Carbon fines from adsorbent attrition in the highly agitated fluidized-bed reactors present a potential particulate emission problem in the exhaust, although control should be possible by moderate pressure-drop cyclones. Removal of NO_x has not been demonstrated, although similar processes such as those of Bergbau-Forschung and Reinluft have reported 40- to 50-percent removal.

The Westvaco process has been demonstrated on a slipstream of 0.16 m^3/sec (330 cfm) from an oil-fired industrial boiler. The system featured three units, a sorber, an acid converter, and a sulfur stripper/hydrogen sulfide generator. The granulated carbon, nominal 12 x 40 mesh size with a density of 641 kg/m^3 (40 lb/ft^3), flowed by gravity between the units. Counter-current gas flows were used in the reactors. The sorbing unit was cooled by direct water spray, and the converter and stripper/generator were heated by electrical resistance. Dust collection was performed by cyclones and bag filters. The system operated for 350 hours, demonstrating better than 90-percent sulfur removal from the inlet concentration of 2000 ppm SO_2.[30] A maximum desulfurization of 97 percent was obtained.[31] There was only one period of downtime, caused by sulfur plugging, and there were no corrosion or dew point problems. The granulated carbon was recycled 21 times with only slight chemical and mechanical carbon loss. Some initial carbon attrition occurred, but it was reduced by using a combination of larger particle sizes with improved hardness. There was no evidence of undersired side reactions or trace contaminant buildup. The carbon had no effect on the 150 ppm of NO_x in the flue gas; it appeared to suppress adsorption SO_2 up to an NO_x concentration of about 150 ppm. The by-product sulfur was of commercial grade, 99.7 percent pure, and contained 380 ppm ash and 2500 ppm carbon.[32] The carbon could be readily filtered, yielding a 99.9[33] percent pure sulfur. A preliminary, 15-MW prototype design has been prepared for 90-percent SO_2 removal from a coal-fired boiler using coal with a 3.5-percent sulfur content.[34]

3.6.4 Bergbau-Forschung/Foster Wheeler Dry Adsorption Process[35,36,37]

The Bergbau-Forschung/Foster Wheeler (BF/FW) process involves the adsorption of dry SO_2 by a moving bed of activated char. The char is then thermally regenerated, and the SO_2-rich gas is reduced to elemental sulfur in a proprietary process using crushed anthracite coal. The system also removes some particulate matter and NO_x from the gas stream. Among the unique features of the process are the louvered, moving-bed adsorber and the use of hot, inert sand to provide heat for char regeneration. Pilot plant studies have indicated 95-percent SO_2, and 40- to 60-percent NO_x removal.[38]

There are three basic process steps in the BF/FW process: adsorption, regeneration, and reduction. The first two were developed by Bergbau-Forschung and the latter by Foster Wheeler. Figure 3-76 is a simplified flow diagram of the process. In the two-stage adsorber, the flue gas passes horizontally in cross flow through vertical columns of activated char. Char in the beds is continuously recycled, moving downward in mass flow and acting as a catalyst for the adsorption of SO_2, SO_3, NO_x, oxygen, and water vapor. Particulate matter is also collected on the char pellet surfaces, which, because of their size and physical arrangement, act as impingement filters. In the second step, the activated char containing sulfuric acid is reduced by being heated to 650°C (1200°F) in an inert atmosphere. Hot sand that is not used in the reaction is used as the heat transfer media. The sand/char mix flows downward through the regenerator to be separated by a vibrating screen deck. The char is spray-cooled to 104°C (220°F) and returned to the adsorber; the sand is recycled to an oil-fired, fluidized-bed heater. Particulate matter is physically separated from

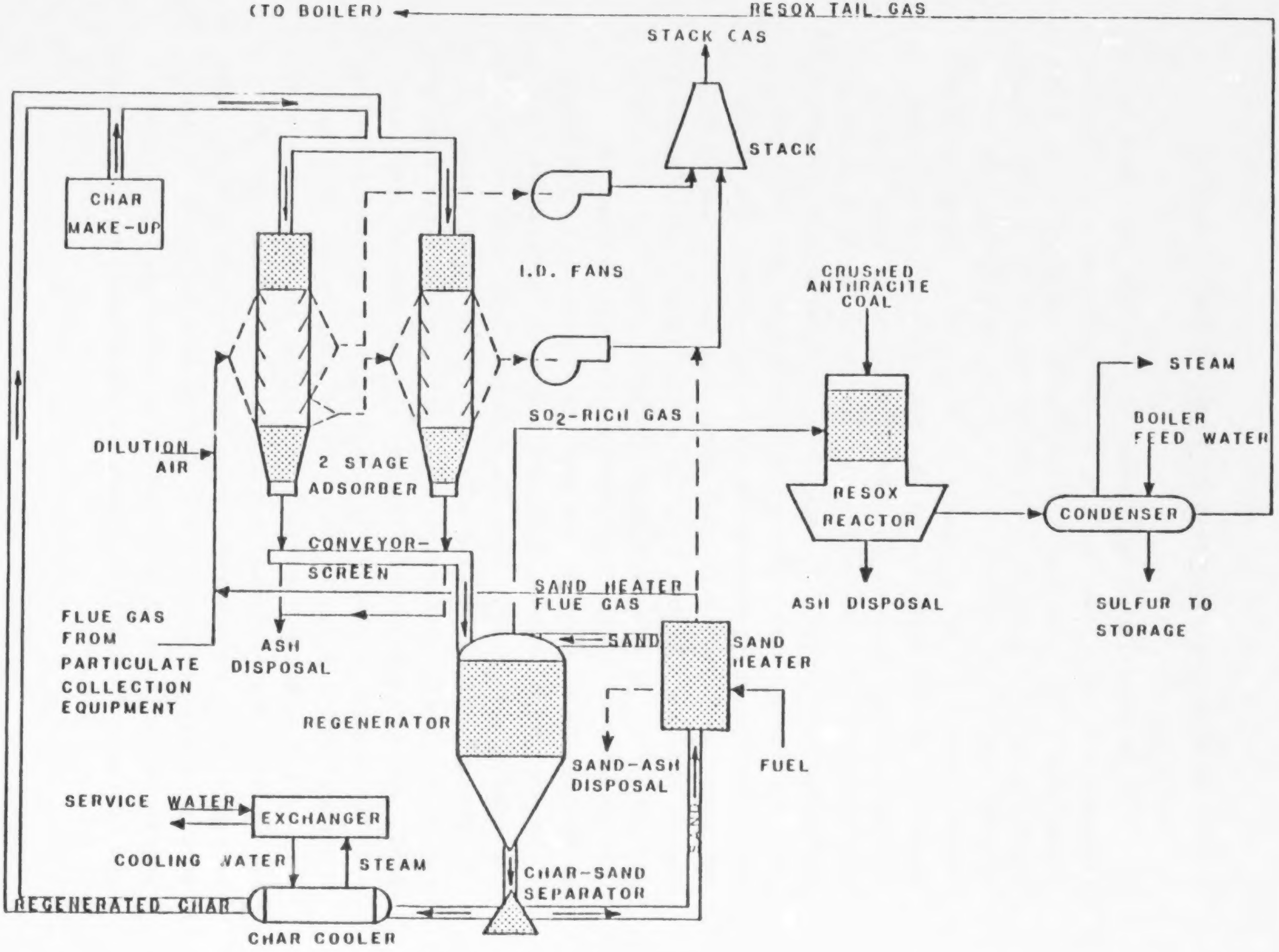

Figure 3-76. Flowsheet - BF/FW dry adsorption process.[39]

the saturated char before regeneration. The offgas from the regenerator contains SO_2 (25 to 40 percent by weight), CO_2, water, and nitrogen. In the third step, Foster Wheeler's RESOX process, the gas stream is introduced in counterflow to a mass flow of crushed anthracite coal, thus reducing the SO_2. Molten sulfur is then recovered in an inclined shell-and-tube condenser, removed to an insulated tank and stored as a liquid. The nitrogen and CO_2 in the gas stream do not take part in the reaction. Tail gas from the condenser is recycled to the boiler to oxidize the remaining sulfur values to SO_2. The chemistry of the BF/FW process is as follows:

$$SO_2 + 1/2O_2 + H_2O \rightarrow H_2SO_4 \qquad (1)$$

$$H_2SO_4 + 1/2C \rightarrow H_2O + 1/2CO_2 + SO_2 \qquad (2)$$

$$2NO + C \rightarrow CO_2 + N_2 \qquad (3)$$

$$SO_2 + C \rightarrow S + CO_2 \qquad (4)$$

As with other dry processes, BF/FW does not require critical chemical control of wet reactions, separation of water from by-products, or stack gas reheating. In addition, the design temperature for the adsorber, 96° to 149°C (205° to 300°F), is about normal for boiler flue gas after an air preheater. In comparison with other FGD systems, BF/FW is intermediate in terms of energy consumption, requiring about 9.5×10^5 J/kWh (900 Btu/kWh) for a 500-MW plant. Other estimated inputs for a plant of this size are: activated char, 1360 kg/hr (3000 lb/hr); electric power, 8 MW; anthracite coal for the RESOX process, 17×10^{10} J/hr (160 MM Btu/hr); the oil-fired sand heater, 22×10^{10} J/hr (210 MM Btu/hr); and total incremental fuel consumption, 9.5×10^5 J/kWh (900 Btu/kWh). Total energy needs are equivalent to approximately 10 percent of the power plant output.

Several potential problems are involved in scaling-up the BF/FW process. Dilution of SO_2 by other gases is a handicap in making sulfur. There is a need for reliable solids-handling equipment that can operate up to 1500°F (the temperature to which the sand is preheated before it is mixed with the char in the regenerator). The materials-handling subsystems for the adsorber and regenerator must either be located near the stack or make use of expensive conveying systems. In some cases this makes retrofit installations difficult. A shortage of activated char could be a problem. High-efficiency ESP's are required to protect the cleaning section from fly ash that has been separated from the char out of the adsorber. An additional waste stream is formed by fly ash, unreacted coal, and tars removed from the bottom of the reduction vessel.

Elements of the BF/FW system have been demonstrated both in Germany and in the United States. The adsorber and regenerator were piloted by Bergbau-Forschung on a flue gas slipstream from the coal-fired boiler at the Kellerman Power Station in Lünen, Germany. A modified Claus plant was used to convert the SO_2-rich gas to elemental sulfur. About 6,000 hours of continuous operation indicated high desulfurization and low pressure drop. Removal of up to 95 percent SO_2, 40 to 60 percent NO_x, and 95 percent inlet particulate was demonstrated.[40] The only effluents were sulfur, cooling water, dry flyash, and small quantities of char fines. The fully integrated BF/FW system has been tested at the Gulf Power Scholz Steam Plant in Florida, where pulverized coal is burned in the boiler. Pilot tests have been conducted on simulated flue gas streams containing various SO_2 concentrations.[41] A high-efficiency ESP removed 99.7 percent of the particulate matter in the gas stream, which contained 900 to 2150 ppm SO_2.[42] Pressure drops in the adsorber were lower than expected, but SO_2 removal was 96 percent or better. Poor char distribution in the adsorber caused imbalances in the bed level and reductions in removal efficiency, and there were also problems in the char feed, char/sand separation, hot sand conveying, and char cooling sections of the regenerator. Two brief runs made with the RESOX reduction system in 1975 indicated that the reaction was controllable. Modifications to the system are planned and testing will continue.

3.6.5 Consol Process[43,44,45]

The Consol process, developed by Consolidation Coal Company, is an aqueous potassium formate process involving SO_2 absorption followed by reduction of the product in solution. Unlike other reduction methods, such as citrate and Shell/UOP, which use sodium compounds for absorption, the Consol technique employs potassium salts. The major product of absorption is potassium thiosulfate ($K_2S_2O_3$), which is reduced in solution with CO at elevated temperature and pressure to produce H_2S. The H_2S can be used for production of elemental sulfur by standard methods. The scrubber has been tested on a 10-MW plant, whereas the regeneration step has been operated only on the laboratory scale. The Consol process is one of three that the EPA has selected for a demonstration project, to be co-sponsored with a large utility.

The chemistry of the Consol process is relatively complex, and many side reactions can occur. Flue gas is first quenched and humidified in a venturi scrubber, which also removes SO_3, HCl, and residual particulate matter (Figure 3-77). The scrubber is a packed tower with a recirculation circuit through a reaction drum. The feed comprises

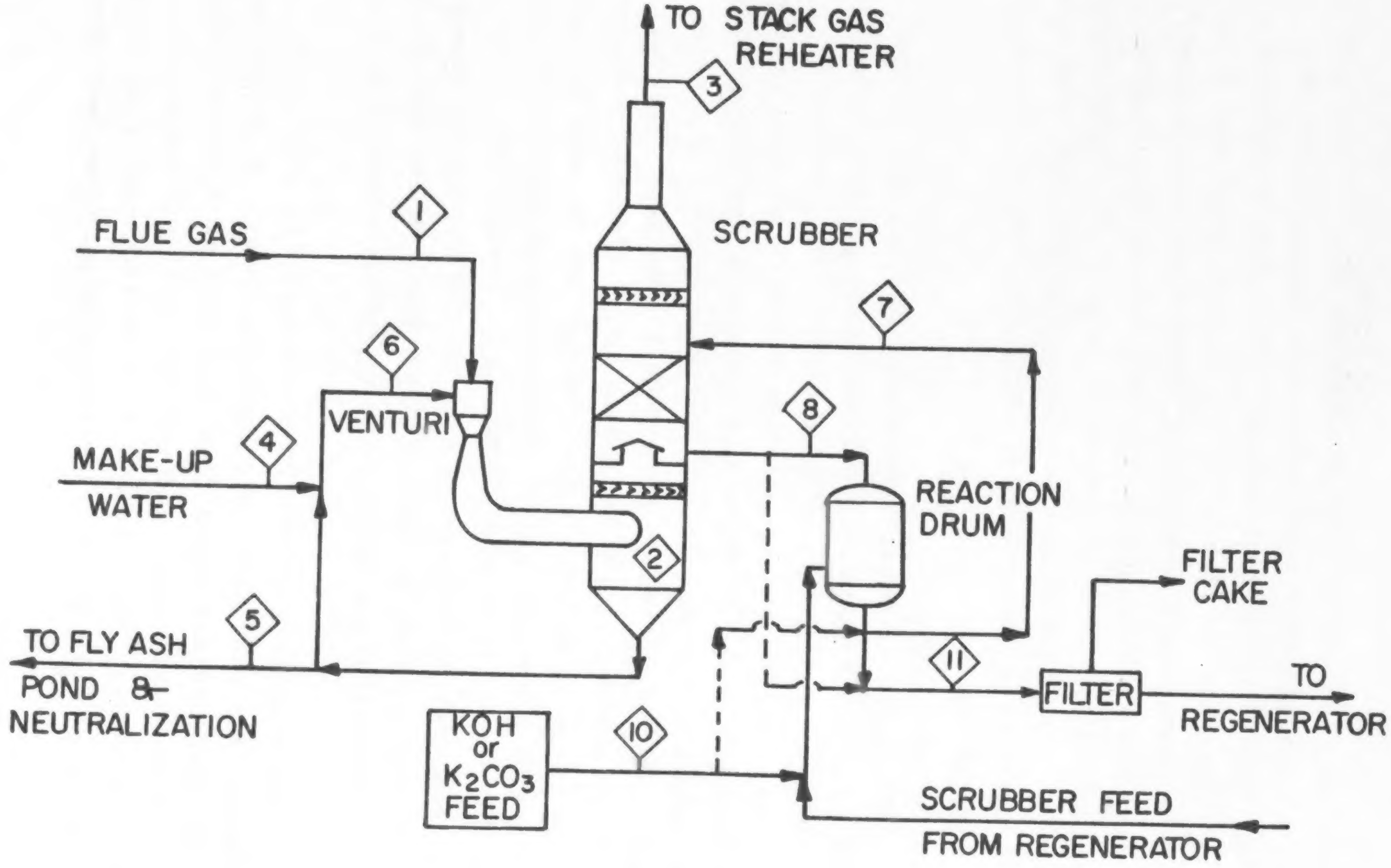

Figure 3-77. Flowsheet - Consol scrubber.[46]

flue gas, spent scrubber product, and makeup potash. Various scrubbing compounds are used, including KHS, K_2S_x, $KHCO_3$, KCO_3, and KOOCH. The key reaction in the scrubber is reduction of potassium sulfite to thiosulfate by the following process:

$$KHS + 2KHSO_3 \rightarrow 3/2\ K_2S_2O_3 + 3/2\ H_2O \qquad (1)$$

The polysulfides are relatively minor components and are consumed in the reaction drum. The KOOCH is a relatively minor component and is essentially inert; it reacts with the circulating scrub solution in the reaction drum. At a pH of 7, a residence time of less than 1.5 minutes is sufficient to consume the K_2S_x and KHS in the reaction drum completely. The KHS must be completely consumed to prevent production of H_2S gas in the scrubber. Sulfur dioxide is removed from the flue gas by reaction with potassium carbonates in the solution, as follows:

$$K_2CO_3 + H_2O + 2SO_2 \rightarrow 2KHSO_3 + CO_2 \qquad (2)$$

The scrubbing process is controlled by adjusting the feed rate to produce an effluent pH of 7, and by maintaining the proper proportions of KHS and K_2S_x to K_2CO_3 and $KHCO_3$.

The basic reaction in the Consol regenerator is the noncatalytic reduction of the thiosulfate scrubbing solution with CO-rich gas:

$$3K_2S_2O_3 + 12CO + 7H_2O \rightarrow 2KHS + 4KHCO_3 + 4H_2S + 8CO_2 \qquad (3)$$

Figure 3-78 is a flow diagram of the regeneration process. Optimum reaction temperature is 232°C (450°F) at a pressure of 3.4 to 6.9 MPa (500 to 1000 psig). The use of a catalyst to maintain the simplicity of the system was tested and rejected. The acceptability ratio is controlled by the proportion of CO_2 in the reducing gas; the ratio of KHS to KHCO3 in the reduction product is controlled by the following equation:

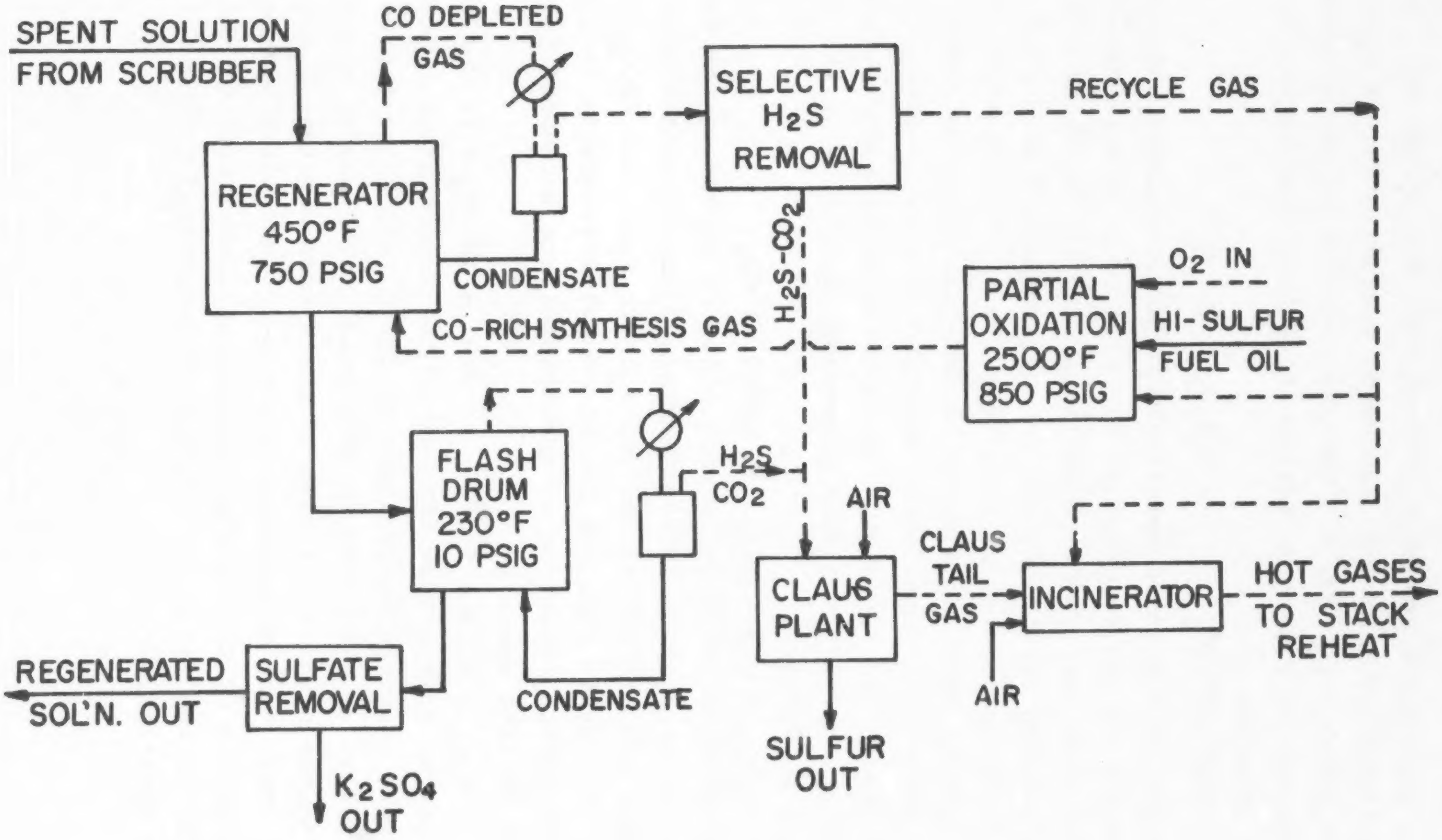

Figure 3-78. Flowsheet - Consol commercial regenerator.[47]

$$KHS + CO_2 + H_2O \rightarrow KHCO_3 + H_2S \qquad (4)$$

Final adjustment of the acceptability ratio to about 0.95 is carried out by decomposition in a flash drum, simultaneously releasing CO_2:

$$KHS + KHCO_3 \rightarrow K_2CO_3 + H_2S \qquad (5)$$

$$2\ KHCO_3 \rightarrow K_2CO_3 + CO_2 + H_2O \qquad (6)$$

Off-gas from the regenerator contains unreacted CO, CO_2, H_2S, and hydrogen. After cooling and steam condensation, the H_2S is selectively removed by scrubbing with a polar organic solvent in one of several available proprietary processes. The absorbed H_2S-CO_2 gas released during regeneration is combined with H_2S-CO_2 gas from the flash drum and sent to a Claus plant for elemental sulfur production. The Claus plant tail gas is incinerated with part of the scrubbed regenerator off-gas to produce hot combustion gas, which is used for direct reheating of the scrubbed stack gas. Carbon monoxide is produced from the regenerator off-gas by a reverse water-gas shift in an oxygen-fired, partial-oxidation unit.

Because of the high solubility of potassium thiosulfate in water, the Consol scrubber can be operated with highly concentrated salt solution. Sulfate problems are relatively minor, because the low sulfite concentration maintained inhibits oxidation to sulfate. Any K_2SO_4 produced is readily marketable.

The Consol scrubber has been demonstrated in a 0.47 m^3/sec (1000-acfm) pilot plant at the Consolidation Coal Research Division in Library, Pennsylvania, and in a 14.2 m^3/sec (30,000-acfm) pilot plant at the Philadelphia Electric Cromby Station. The Library unit received feed-gas from an oil-fired boiler; the Cromby installation was on a slipstream from a coal-fired boiler. The regenerator has also been operated at Library, including a closed-loop integrated operation with the scrubber. Sulfur dioxide absorption at Cromby varied from 88 to 99 percent, with the higher rates achieved by improved liquid distribution, smaller packing, and a higher gas rate. Sulfate formation averaged about 1 percent of the SO_2 absorbed. There was no appreciable NO_x absorption. In a 10-day run of the closed-loop, integrated operation, sulfur absorption averaged about 99 percent, with no accumulation of foreign material in the product.

3.6.6 Aqueous Carbonate Process[48,49,50,51]

In the Aqueous Carbonate Process, developed by Atomics International, SO_2 in the flue gas is absorbed by a very concentrated sodium carbonate solution in a spray dryer. The finely-divided sodium sulfite powder that results is reduced with coke to regenerate the carbonate and to produce hydrogen slufide. This can be used for the production of elemental sulfur by standard methods. The process has been designed for the removal of 90 to 95 percent of SO_2 from the flue gases of coal-fired power plants. The carbonate solution is said to provide effective SO_2 removal over wide concentration ranges.[52] In addition to its use of a spray dryer for producing a granular salt mixture suitable for regeneration, the process is also unique in that there is complete reduction of the sodium salts in a molten pool. There is no need for reheat, since the flue gas is not saturated during contact with the solution. Individual components of the system have been pilot tested, and the process is also one of the three chosen for the EPA demonstration project.

There are six major process steps in the Atomics International system: scrubbing, product collection, reduction, quenching and filtration, carbonation, and sulfur production (Figure 3-79). The flue gas is precleaned to 0.22 g/m^3 (0.1 gr/scf) flyash in cyclones or in an ESP. The spray dryer is similar to those used in other applications for more than 50 years. Sodium carbonate solution is atomized by high-speed centrifugal atomizers, and the fine mist absorbs SO_2 in crossflow. The thermal energy of the water gas vaporizes water without excessive saturation or gas cooling. Solid sodium sulfite and sulfate granules are formed and removed from the exit gas in cyclones or in an ESP. Because the waste gas remains above the dew point, it can be vented directly and without reheat. The reduction unit is a refractory-lined, steel vessel which contains molten sodium carbonate and sodium sulfide at 930° to 1040°C (1700° to 1900°F). Dry feed from the spray dryer melts and mixes with the pool, and the sulfite and sulfate are reduced by petroleum coke to sulfide. Off-gas from the reducers is used as a source of CO_2 for later carbonation steps and as a source of process heat. The molten mixture flows to a quench tank, where it is dissolved. The solution is then cooled and filtered to remove carbon and flyash. The quench and filter units are similar to those used in the pulp and paper industry. Quenching can be eliminated by cooling the reduced melt, allowing it to solidify, and breaking it into chunks. The solution is precarbonated with pure CO_2 to produce sodium hydrosulfide. Final carbonation products are rich in H_2S and are combined and sent to a Claus plant. The slurry from the bicarbonator/crystallizer is decomposed to produce sodium carbonate solution for the spray dryer and pure CO_2 for the precarbonator. The principal chemical reactions occurring in the Aqueous Carbonate Process are as follows:

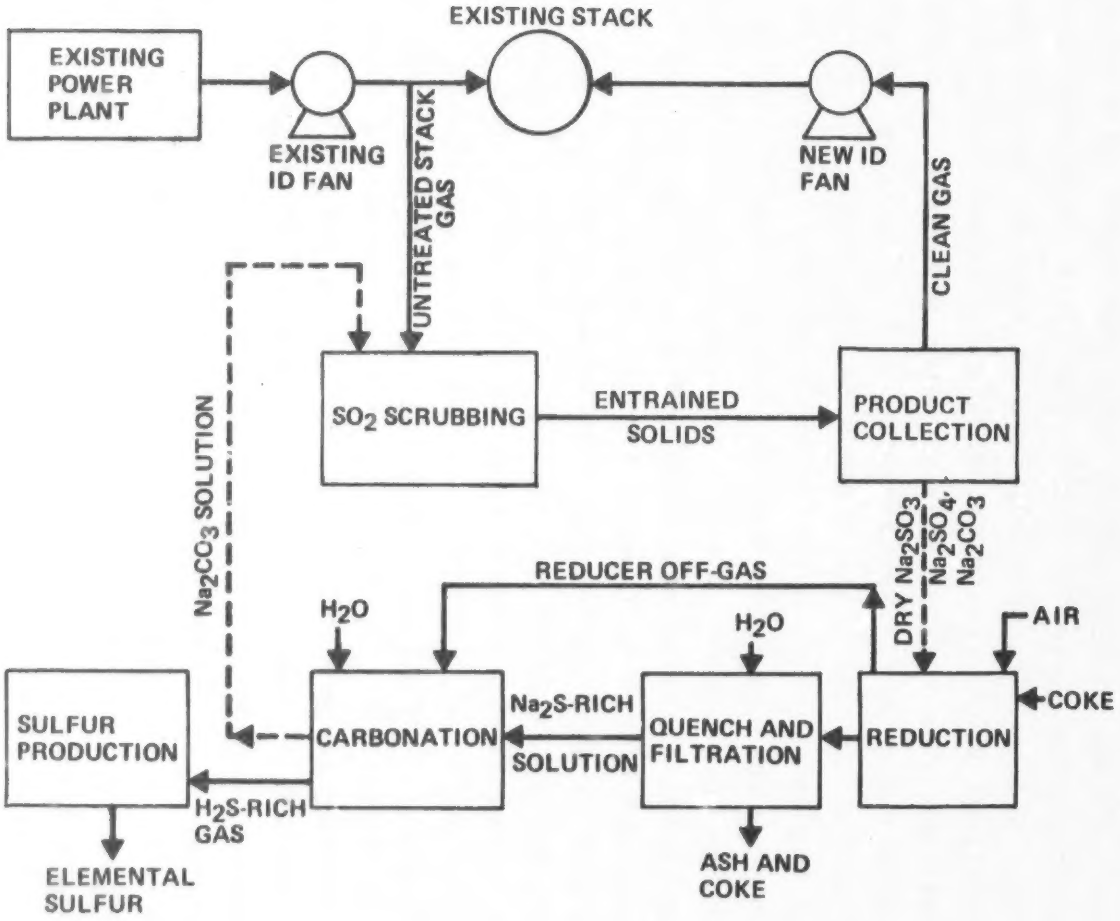

Figure 3-79. Flowsheet - Aqueous Carbonate Process.[53]

$$SO_2 + Na_2CO_3 \rightarrow Na_2SO_3 + CO_2 \quad (1)$$

$$Na_2SO_3 + 3/2\ C \rightarrow Na_2S + 3/2\ CO_2 \quad (2)$$

$$Na_2SO_4 + 2C \rightarrow Na_2S + 2CO_2 \quad (3)$$

$$Na_2S + CO_2 + H_2O \rightarrow Na_2CO_3 + H_2S \quad (4)$$

In addition to the fact that reheat is not necessary, there are several other advantages to the Aqueous Carbonate Porcess. Petroleum coke is not only an economical reducing agent; it reduces all sulfates, and thus avoids water pollution problems. Coal would be a desirable reducing agent, but it produces problems with flyash, chlorides, and trace contaminants. Although the Aqueous Carbonate process has not been tested on a full-scale application, similarly-sized dryers are being used successfully on other processes. A marketable sulfur by-product is produced, and no corrosion problems have been reported. In addition to SO_2 removal, there are also significant reductions in SO_3, particulate and halogen gases. Removal of NO_x was found to be less than 5 percent, based on normal pilot plant operating conditions. The only waste streams consist of a flyash filter cake discharge and the chloride purge stream discharge contained in the spent cooling water. The process is a relatively efficient energy user. Based on a design application to a 500-MW plant, total incremental fuel consumption is 72×10^4 J/hr (680 Btu/kWh). Other requirements for a plant of this size include: reducing coke, 21×10^{10} J/hr (201 MM Btu/hr); fuel oil, 32×10^9 J/hr (32 MM Btu/hr); electricity, 10 MW; and sodium carbonate, 385 Kg/hr (850 lb/hr).

The relatively large size of the Aqueous Carbonate System, especially of the spray dryer, may cause problems in retrofit installations. A particulate collection system is necessary to recover sodium sulfite from the dryer off-gas, and a secondary particulate collection system will be needed. Other potential problems include relatively high heat requirements in the reduction step, control of the molten salt reactor, material heat and corrosion stress, and the possibility of a quench-melt explosion.

The individual process steps have been separately tested. Spray dryers were pilot tested in a unit with a diameter of 1.5 m (5 ft) on a slipstream from a coal-fired power plant at Southern California Edison's Mohave Station, and in a unit with a diameter of 2.1 m (7 ft) at Bowen Engineering's North Branch, New Jersey, laboratory. Over 100 tests were conducted at Mohave, with the dryer and cyclone operating up to 1.9 m^3/sec (4000 scfm). Over 90 percent SO_2 removal was achieved on a flue gas containing from 200 to 4000 ppm SO_2.[54] The cyclone collectors removed from 89 to 99 percent of the sodium sulfite and sulfate product.[55] At the Bowen laboratory, it was found that SO_2 removal efficiency could be maintained by adjusting the solution feed rate in direct proportion to the feed gas rate. The regenerator has been tested in the Atomics International laboratory. Integrated tests are still needed to establish the viability of this process.

3.6.7 Aqueous Sodium Carbonate Absorption[56,57,58]

A throwaway aqueous FGD system has been developed by Combustion Equipment Associates. Trona, a hydrous acid sodium carbonate ($Na_2CO_3 \cdot NaHCO_3 \cdot 2H_2O$), is powdered, slurried, and used as the absorbent in venturi-type scrubbers. Various alternate forms of sodium carbonate have also been employed in the system, which has been successfully tested in both pilot and commercial scale on a coal-fired power plant.

Many of the design details of the trona scrubbing system are the property of Combustion Equipment Associates, but the commercial installation at Nevada Power's Reid Gardner Station in Moapa, Nevada, can be described. A separate modular system, consisting of two parallel venturi scrubbers followed by a single-stage perforated-plate washing tower, has been installed on each of two 120-MW boilers burning pulverized coal. The design throughput of each scrubber is 223 m^3/sec (473,000 acfm) at 177°F (350°F). Seventy-five percent of the flyash is first removed from the hot flue gas in a multicyclone; the gas is then split and sent to the twin scrubbers, where it is quenched with scrubber liquor sprayed from tangential nozzles.[59] The pressure drop across the scrubbers is 3.7 kPa (15 in. H_2O). The gas then goes to a cylindrical separation tower, in which finer droplets are removed by centrifugal force. It is bubbled through a sieve tray and passed through a horizontal, single-pass, radial-vane-type mist eliminator. Prior to discharge, the gas is reheated by direct mixing with hot air that has been indirectly heated with steam in a heat exchanger. The scrubber liquor is prepared by mixing trona with water from the ash pond to dissolve the sodium carbonate. Insolubles, mainly sand, are settled from the slurry in a clarifier. The scrubber liquor enters a recirculation circuit, in which the blowdown from an intial tank is mixed with the alkaline clarifier underflow and neutralized in a tank to pH 7. Spent liquor from this tank, about 1 percent of the recycle stream, is pumped to an ash settling pond and the overflow to a larger evaporation pond. The sieve tray is flooded with water from the ash pond, which is also the source of the water mixed with the recycle liquor stream before it is returned to the venturi. The chemistry of the trona system is based on the following reactions.

$$Na_2CO_3 + SO_2 \rightarrow Na_2SO_3 + CO_2 \qquad (1)$$

$$Na_2SO_3 + SO_2 + H_2O \rightarrow 2NaHSO_3 \qquad (2)$$

An open-loop soluble system such as the aqueous sodium carbonate process minimizes operating difficulties that are due to plugging and scaling. In addition, there is no problem with absorbent volatilization. Disposal of soluble sodium compounds, however, may create water pollution problems.

The trona system was piloted in 1972 with a unit of 3.3 to 4.2 m^3/sec (7000 to 9000 acfm) capacity at the Reid Gardner Station. The full-sized units began operation in April 1974. The sulfur content of the coal is between 0.4 and 0.7 percent, averaging 0.55 percent. No major operating problems have been reported, and the scrubbers have achieved greater than 90 percent SO_2 removal efficiency (326 to 577 ppm inlet and 1 to 26 ppm outlet).[60] The evaporation pond is lined with clay. Because of the low relative humidity in the area, this is thought to be appropriate for the disposal of spent liquor. A shortage of trona, however, prevents widespread use of the aqueous sodium carbonate system.

3.6.8 Shell/UOP Copper Oxide Adsorption[61,62,63]

The Shell/Universal Oil Products process is a dry metal oxide system, using copper oxide (CuO) on an activated sorbent. The copper oxide is regenerated, and the SO_2 released for further processing to elemental sulfur or liquid SO_2. A dry, selected adsorbent is used so that the complications of a wet system can be avoided. Copper oxide was chosen because of its high reactivity with SO_2, and because it can be regenerated without thermal energy. A unique feature of the process is its use of parallel passage, fixed-bed reactors. The Shell/UOP process has been installed on a commercial-sized, oil-fired boiler in Japan. It, too, is one of the three processes chosen by EPA for its demonstration project; a demonstration unit of this type has been built on a slipstream from a coal-fired utility boiler in Florida.

The Shell/UOP process features simultaneous adsorption/regeneration, followed by SO_2 recovery and concentration (Figure 3-80). A fixed, packed bed is used for adsorption, since moving beds would involve greater attrition of the copper oxide acceptor, increased particulate separation, and relatively complex solids-handling operations. The gas stream flows through open channels beside, and in contact with, the copper oxide. This results in a low pressure drop and reduced particulate plugging. As the acceptor becomes loaded with SO_2, collection efficiency decreases and copper oxide is regenerated in-situ at the same temperature 400°C (750°F). Reducing agents such as hydrogen, carbon monoxide, or light hydrocarbons are used for regeneration; hydrogen is preferred because of its high reactivity with copper sulfate. The rate of SO_2 release can be controlled by the rate of hydrogen addition. Hydrogen is catalytically oxidized by copper oxide to water; the SO_2, free of oxygen and particulate matter, can be used for the production of liquid SO_2 or elemental sulfur. The use of two or more reactors in parallel, with some in service while others are being regenerated, permits continuous SO_2 removal. There are a number of ways to concentrate and prepare SO_2-rich regeneration gases for further processing, including water adsorption/stripping, solvent absorption, compression and condensation of water vapor, and complete liquefaction. In addition, UOP has developed other less energy-intensive procedures. The basic chemistry of the Shell/UOP process is as follows:

$$CuO + 1/2O_2 + SO_2 \rightarrow CuSO_4 \quad (1)$$

$$CuSO_4 + 2H_2 \rightarrow Cu + SO_2 + H_2O \quad (2)$$

$$CuO + H_2 \rightarrow Cu + H_2O \quad (3)$$

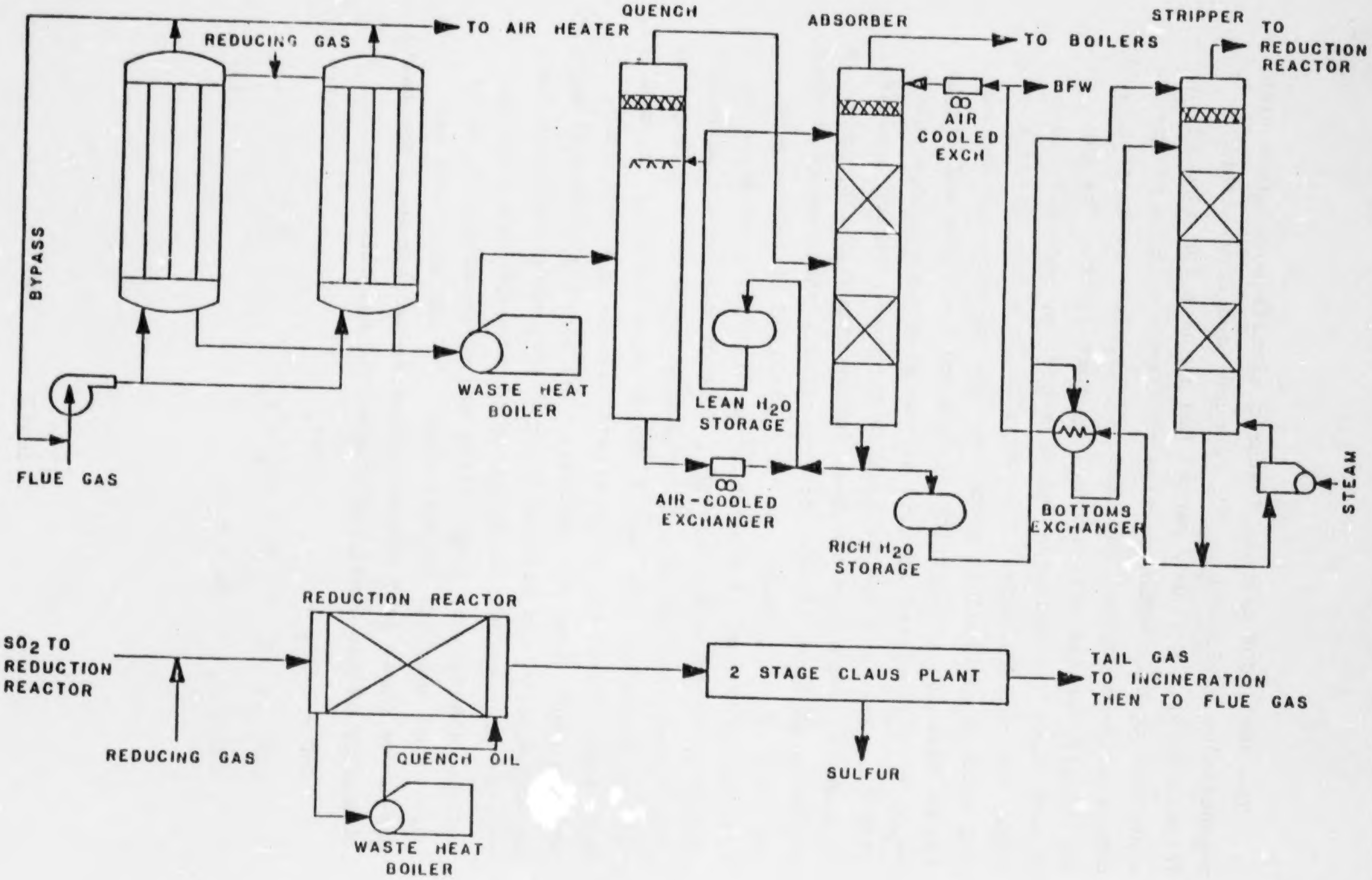

Figure 3-80. Flowsheet - Shell/UOP copper oxide adsorption process.[64]

UOP guarantees 90-percent SO_2 removal with the copper oxide adsorption process.[65] In addition, the process provides almost complete SO_3 removal. The only waste stream is wastewater, containing about 50 ppm SO_2 at 455 to 680 lpm (120 to 180 gpm) for a 500-MW plant. Removal of 60 to 70 percent of NO_x has been demonstrated by the addition of ammonia.

Integration of a complete copper oxide system into a power plant will be difficult, because a cyclic fixed-bed adsorber and a variable flue gas stream cannot be compatibly operated in conjunction with relatively inflexible hydrogen and sulfur production units. An inlet dust load of less than 0.22 g/m^3 (0.1 gr/scf) is required. The fairly large space requirements, and the necessity of reheating the exit gas from 150° to 400°C (300° to 750°F), could create difficulties in some retrofit installations. In addition to the large amount of hydrogen reducing gas needed, energy demands are high, because of the quantity of water circulated and the steam necessary to heat the stripper reboiler. Total incremental fuel requirements for a 500-MW plant are 17×10^5 J/kWh (1620 Btu/kWh); other requirements include: ammonia, (2110 lb/hr); reducing agent, 34×10^{10} J/hr (322 MM Btu/hr); steam, 30×10^{10} J/hr (288 MM Btu/hr); fuel oil, 52×10^9 (50 MM Btu/hr); electricity, 6 MW.

The Shell/UOP acceptor and regenerator were first demonstrated in 1967 on a slipstream of 0.19 to 0.28 m^3/sec (400 to 600 scf/min) from a process heater using high-sulfur fuel oil at Shell's Pernis refinery near Rotterdam. Ninety-percent SO_2 removal was demonstrated over 20,000 operating hours, with estimated acceptor life of 1.5 years.[66] A commercial unit was installed in 1973 at a Showa Yokkaichi Sekiyu (SYS) oil-fired, 40-MW boiler (125,000 Nm^3/hr - 2500 ppm SO_2).

Only the acceptor and water-gathering systems were installed, because there were existing hydrogen and Claus plants. This system had 68 unit-cells, identical in design but larger than the single cell at Pernis. Ninety-percent SO_2 removal was shown, with the SO_2 co-processed both in the Claus plant with normal refinery acid gas, and in an absorber/stripper. The longest continuous run was 2 months.[39] Simultaneous removal of 60 to 70 percent NO_x was achieved by the addition of ammonia, with a maximum ammonia carryover in the exit gas of 2 ppm. Hydrogen consumption was 0.2 percent by weight of sulfur recovered.[67] The stripped water contained 20 ppm sulfur, 75 percent of which was present as sulfate.[68] A demonstration unit of the acceptor and regenerator has also been installed on a slipstream from a coal-fired boiler at Tampa Electric's Big Bend Station. It has five modules identical to those at the SYS refinery, and is designed for 90-percent desulfurization of a gas sgream with 1260 to 1400 ppm SO_2. Operating results have not been reported. The EPA-utility joint demonstration plant will be the first completely integrated unit on a coal-fired boiler.

3.6.9 Alkaline Flyash Scrubbing Process[69,70]

In the alkaline flyash scrubbing process, the inherent alkalinity of the flyash from certain coals is used to react with SO_2. Lime is added to the scrubbing liquor for pH control. The alkaline species in the scrubbing liquor combine with the SO_2 to form insoluble sulfates and sulfites, which are then disposed of. The system has been successfully tested in four pilot plant studies, and is currently operated on two utility boilers. It will be employed on six utility boilers in the future.

Lime/alkaline flyash scrubbing is a throwaway process, very similar to lime systems, except that the major source of the active alkali is flyash from coal. It is used primarily in the West, where the flyash of some coals is high in sodium oxide, calcium oxide, and magnesium oxide. The alkali content of Western coal ash varies from under 10 to over 50 percent, with the lower-rank coals (i.e., lignite) having the highest alkali content. Figure 3-81 is a simplified flow diagram for a lime/alkaline flyash scrubbing system. Flue gases enter the scrubber at approximately 150°C (300°F). The scrubber applied to large utility boilers is a two-stage device with a venturi and a spray tower in series. In the venturi, most of the flyash is collected, whereas the spray tower absorbs SO_2. Slurry from both stages is collected and held in the bottom of the scrubber for reuse. The slurry is usually agitated to hold the flyash in suspension. The cleaned flue gas passes through a mist eliminator and reheater before discharge to the stack.

A purge stream is constantly withdrawn from the recycle slurry in order to maintain approximately 12 percent solids in the recycle stream. The purge stream is usually clarified or thickened before disposal in a pond. Decanted water from the clarifier/thickener is recycled to the absorbent loop.

Makeup water is also added to the absorbent loop, to compensate for water losses by evaporation and by retention in sludge. The scrubber recycle is maintained at a constant pH of about 5.0 by the addition of lime or limestone slurry to the recycle stream. This is done to minimize scaling and to promote ionization of the active alkali species in the flyash.

Based on operation experience, the lime/alkaline flyash system is considered attractive. Tests on Colstrip Units 1 and 2 revealed that an emission level of 0.19 kg/10^6 J (0.44

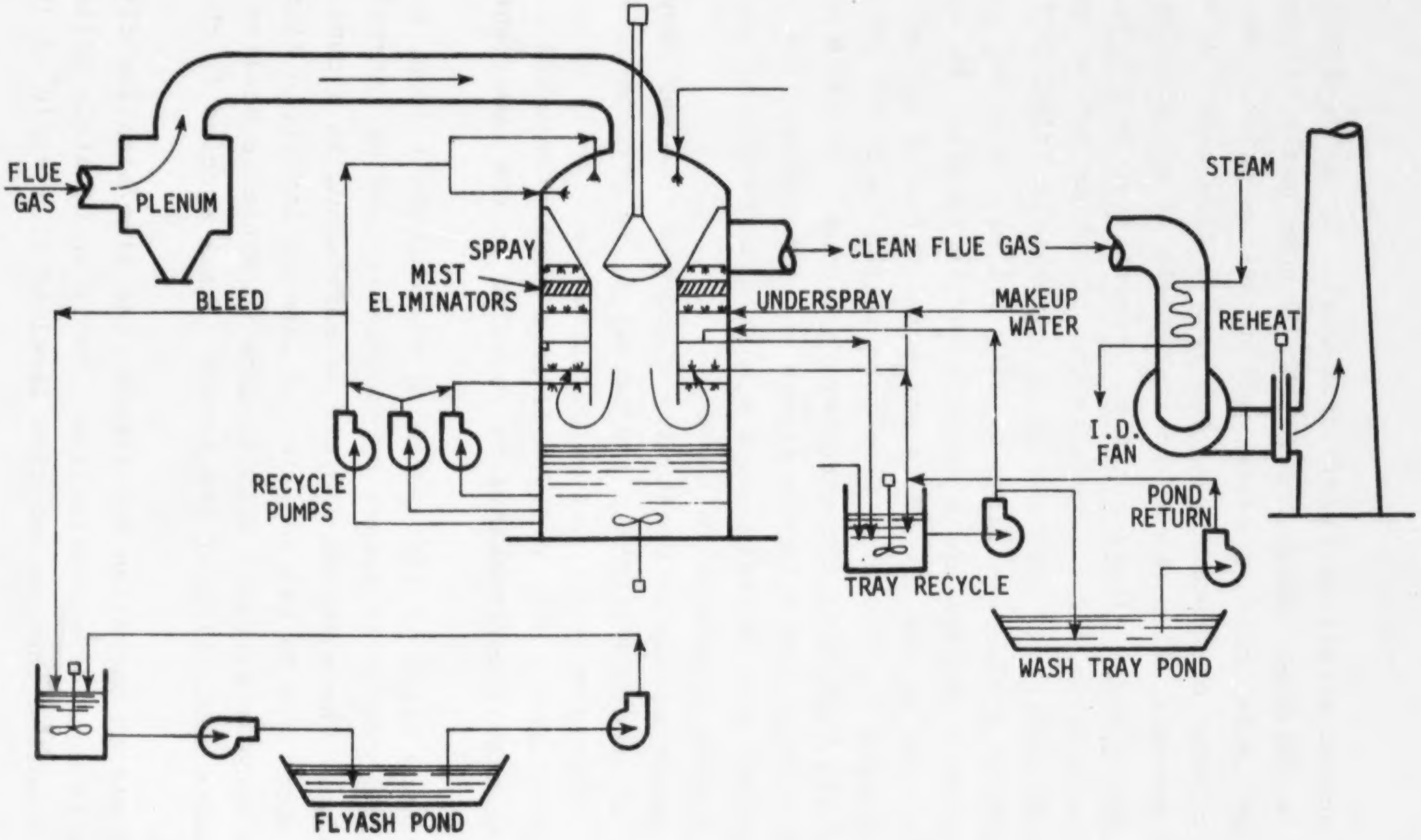

Figure 3-81. Flowsheet of lime/alkaline flyash scrubbing.[71]

lb $SO_2/10^6$ Btu) or less could be attained using coal with sulfur contents of 0.56 to 0.94 percent.[72] No measurements were made of inlet SO_2 loadings to the scrubber; however, assuming a 90-percent oxidation of the coal sulfur to SO_2 in order to estimate the inlet SO_2 loadings, an efficiency of 70 to 90 percent was achieved. The major limitation to the use of alkaline flyash scrubbing is the need for high active alkali concentrations in the ash in relation to the sulfur content of the fuel. Application would therefore be limited in the United States to low-sulfur, high-alkaline Western coals.

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4.0 PERFORMANCE AND OPERABILITY OF FGD SYSTEMS

This section summarizes the current status of FGD technology with respect to SO_2 removal efficiency and system operability. All of the documentation in this section is presented by reference to specific full scale or test facilities. Information for these plants is presented in detail in Section 3 and the appendicies of this report. This information was obtained from the extensive research that has been reported for pilot and demonstration scale units in this country and also included data from full scale facilities. The design and operating experience gained with first generation FGD systems has resulted, and will continue to result, in improved design and operation of subsequent installations. FGD systems which are currently being engineered will incorporate the latest design features and even better performance can be expected as newer plants are completed.

New plants that are designed with current knowledge of process kinetics and chemistry, equipment capabilities, experience with materials of construction, and adequate redundancy can achieve removal efficiencies and operabilities of, or in excess of, 90 percent.

In this section of our report we have summarized the operating experience and problems which FGD systems have encountered; the solutions (both engineering and operational), which have resulted; and the operability, availability and control efficiencies that can be reasonably expected in FGD technology.

4.1 OPERATING PROBLEMS AND SOLUTIONS[1,2]

There have been and still are problems associated with FGD systems; however, the majority of these problems have basically been solved and the methods of reducing the severity of the remaining items are much better understood. While these problems can be minimized, it must be recognized that large complex chemical processes like an FGD system need constant attention and maintenance, and will never be run in a completely troublefree manner.

To date, the problems encountered with FGD systems and the severity of these problems varied both with system type and within units of the same system. The more common problems encountered are listed below, followed by a more detailed discription of each problem and how it is being solved.

- Formation of scale in the absorber and associated equipment in lime and limestone systems leading to plugging and reduced capacity
- Plugging of mist eliminators, lines, and some types of absorbers
- Failure of ancillary equipment such as pumps, piping, pH sensing equipment, reheaters, centrifuges, fans and duct and stack linings
- Inadequate absorbent make-up preparation
- Handling and disposal of sludge in nonregenerable systems

4.1.1 Scaling and Plugging

In lime and limestone systems, scaling has been a particular problem and has reduced operability in a number of systems. Both a soft sulfite scale and a hard sulfate scale may form in the absorber, mist eliminator, and ancillary tanks, pumps, and pipes. Much effort and investigation has shown that these problems are closely related to the system chemistry and therefore, specific process control techniques have produced significant improvements:

° Operation at subsaturation levels for calcium sulfate and sulfite.

Full scale and test facilities in this country have effectively reduced saturation and scaling by addition of magnesium to the circulating slurry. The TVA Shawnee facility, the Phillips facility, and the Paddy's Run facility demonstrated that the addition of magnesium to the lime and limestone slurry eliminated scrubber scale formation. Other large stations, the Bruce Mansfield and Conesville use lime containing magnesium oxide to prevent scaling.

By maintaining high liquid to gas (L/G) ratios, the proportion of unreacted lime or limestone remains high relative to the absorbed SO_2. There is thus less chance of creating a supersaturated solution of sulfites or sulfates. The higher L/G ratio also improves overall SO_2 collection efficiencies as described in Section 3.1. The actual L/G will vary with the type of absorber, and values in excess of 10.8 l/m^3 (80 gal/acf) have been used in spray towers.

Increased reaction tank holding time will also decrease saturation by allowing further reaction between the absorbed SO_2 and the lime or limestone slurry. Slurry residence time at the Green River facility is greater than twenty minutes and scale formation is not a major problem. This plant has an excellent availability history.

° pH Control.

Work at the EPA-Shawnee test facility has shown that the most important parameter in controlling scale formation is solution pH. Operational experience has shown that lower pH values accelerate scale formation. This is documented at full scale facilities such as LaCygne. The measurement of pH has also received considerable attention. More rugged and dependable sensors are being used; they are located in the slurry stream where they are subject to less breakage, are more accessible, and where they yield data which is more reliable and responsive for pH control. The Bruce Mansfield facility has just completed a renovation program to incorporate these design features into their pH system.

° Co-precipitation of sulfate.

Minimizing the oxygen content in the flue gas, by reducing any air in-leakage, favors co-precipitation of sulfate with the sulfite crystal. Therefore, air exposure is reduced by covering open reaction tanks, clarifiers, etc.

Plugging caused by deposition of solids on equipment surfaces has sometimes restricted the passage of liquids or gas in FGD systems. It is usually easily removed by flushing with water or steam. Plugging in pipes can be prevented through designs which avoid low flow velocities. Careful control of raw material particle size and screening of the slurry also decreases plugging problems, especially in spray nozzles, pipes and pumps. Since this problem is caused by the deposition of solids from the recirculating slurry, reduction of the overall amount of solids will reduce the plugging. The minimum stoichiometry that will effect the required SO_2 removal efficiency should be used. This has been demonstrated in this country at Shawnee and LaCygne. The Japanese facilities use stoichiometries in the 0.95 to 1.05 range, have no plugging difficulties and have SO_2 removal efficiencies in excess of 95 percent.

In summary, the scaling/plugging problem can be controlled or eliminated by carefull scrubber chemistry control. This has been demonstrated at operating facilities in this country and Japan.

4.1.2 Erosion and Corrosion

Problems with ancillary equipment plagued many of the first FGD systems because of poor design, faulty component manufacture, or improper installation. Many of these problems were due to corrosion and erosion.

Erosion in venturi pre-scrubbers has resulted from high fly-ash loadings. Likewise, pre-scrubbers remove the bulk of any chlorides and sulfur trioxide in the gas stream; both of these components are highly corrosive. Corrosion occurs more frequently in areas subject to wet saturated flue gas as opposed to areas subject to alkaline slurry streams.

There are so many factors involved in FGD operation which effect corrosion rates, that generalizations regarding corrosion resistant materials is difficult. A sufficient amount of data has been accumulated, however, to provide general guidelines for the construction of critical elements in FGD systems as summarized below:

(a) Some systems are incorporating such alloys as Hastelloy C-276, Hastelloy G, Inconel 625, Incoloy 825, 317L stainless steel, 904L stainless steel and Jessop JS700 in wet/dry high temperature, high chloride environments, such as in presaturators. The LaCygne Station has found that these materials give excellent reheat service. The Bruce Mansfield station has had good results with Hastelloy wetted parts in the fan.

(b) Synthetic and natural rubber coatings predominate in recycle tanks, pumps, and lines. These materials have been reported to give superior erosion resistance once application problems have been overcome. For instance rubber lined pumps have been used successfully at the following facilities: Green River, LaCygne, Bruce Mansfield, and Conesville.

(c) For liners in the absorbers, exhaust ducts and stacks, a number of materials such as resins, ceramics, polyesters, polyvinyls, polyurethanes, carboline, and Gunite, have been used with varying degrees of success. Although successful applications have been reported, widespread failures of the liners have been attributed to the undependability and inexperience of lining applicators, instability of the materials at high temperatures, inconvenience of repair, and cost-related factors. These problems are especially evident on higher sulfur coals. Extensive effort in continuing by FGD suppliers to fully solve this problem.

In summary, the corrosion and erosion problems have been solved by employing more expensive materials of construction. It is apparent that the technology exists to control these problems, but it is more expensive than materials that are normally used.

Increased SO_2 removal efficiency should have no additional effect on corrosion or erosion problem in systems which are designed properly. In fact, since higher SO_2 removal efficiencies will reduce the SO_2 content in exit gases the downstream equipment corrosion may be reduced.

4.1.3 Equipment Design Improvements

Solutions utilized to reduce problems with ancillary equipment include:

° Recirculation Pumps - Slurry recirculation pumps provide the driving force for the liquid circuit in FGD systems. In their design, special attention must be given to an accurate service description (solution pH, specific gravity, solids content, gas entrainment, flow rates, and head). A number of general trends are evident, and summarized below:

(a) New systems must incorporate spare pumps. Spare capacity from 50 percent (one spare for every two operational) to 100 percent (one spare/one operational) is useful to avoid downtime.

This type of spare equipment is found at many new large stations including Bruce Mansfield and Conesville.

(b) Natural and synthetic molded rubber lining should be specified for wetted parts in the pumps.

(c) Flush-water wash systems are needed to purge the pumps of solids, which tend to settle out during periods of inactivity.

° Mist Elimination - Chevron and baffle-type mist eliminators have been and are currently being used in virtually every FGD system in the United States. The popularity of these collectors is due primarily to design simplicity, high collection efficiency (for moderate to large size drops), low pressure drop, wide-open construction, and low cost. Within these two preferred types of mist eliminators, a number of specific design and construction innovations have been implemented:

(a) Chevron designs (continuous vane construction) are predominant over baffle designs (discontinuous slat construction).

(b) Fiberglass-reinforced plastic is now used at nearly all facilities.

(c) The horizontal configuration (vertical gas flow) is also used in almost all installations for cost reasons.

(d) Two-stage designs predominate over single-stage designs, because they yielded higher elimination efficiencies.

(e) Bulk entrainment separators, perforated plates, impingement plates and other precollection devices are becoming integral parts of mist elimination systems. These reduce plugging and improve separation. The Conesville facility employs this as well as LaCygne, Lawrence, Cane Run No. 5, and Coal Creek.

(f) Mist eliminator wash systems that employ intermittent, high-velocity sprays predominate over continuous wash systems. These produce a hydraulic washing effect.

(g) Operation at high alkali utilization.

> Mist elimination in streams containing slurry droplets is not new technology. The pulp and paper and sulfuric acid industries have been using it for many years. The problems that are encountered today in the FGD industry are similar to those that have been experienced in other industries for some time. New design and operating conditions are continually reducing the problems. For instance, Combustion Engineering offers a bulk entrainment separator system that has worked effectively at the Lawrence and Sherburne stations.

° Reheat - Virtually all the FGD systems coming on-line and planned for future operation incorporate some type of stack gas reheat system. These systems heat the flue gas to avoid condensation with subsequent corrosion to downstream equipment, ductwork and stack and to suppress plume visibility as well as enhance plume rise and pollutant dispersion. To date, a number of "wet stack" FGD systems (no reheat) have been installed and have encountered corrosion problems. The trend in reheat systems is towards heating of ambient air and mixing with the flue gas and mixing of untreated flue gas with scrubbed gas.* In-line reheat systems have been subject to corrosion and solids deposition, the latter often occurring because of inefficient upstream mist elimination.

Several existing facilities have found that they could control reheater problems by fabricating the reheat bundles from more expensive materials. Both Will County and Sherburne have found this to be true. At Sherburne and Lawrence, carbon steel fin-heaters have been used with hot water as the heating mechanism. Reheat bundle life expectancy is 5 to 7 years, which is considered acceptable for this service. Since these tubes are carbon steel, corrosion failure is not catastrophic but gradual. This permits planned shutdown and maintenance. Both of these units have excellent mist elimination systems (multiple-stage FRP mist eliminators which incorporate bulk entrainment separators), which contribute to the reheat success.

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* Blending with untreated flue gas is, of course, limited when high efficiency is desired. The amount of gas which can be bypassed for reheat purposes depends upon the amount of SO_2 resulting from the coal, the applicable regulations, the particular emission levels.

° Fans - Fans installed immediately after an FGD system (wet fans) have experienced corrosion, chloride attack, and solids deposition problems. Deposition problems have caused fan imbalance resulting in excessive bearing wear and damage to the fan. Three systems have reported this problem; Phillips Bruce Mansfield, and St. Clair. The problems associated with fans installed upstream of the FGD system (dry fans) include operation at higher temperatures (over 150°C) resulting in higher gas velocities and abrasion by fly ash. Dry fan problems are more easily solved, and the tendency is toward fans upstream of the FGD system. Where necessary, however, the use of various steel alloys have made wet fans a viable alternative, as at Phillips and Bruce Mansfield.

° Summary - It has been demonstrated that each of the major equipment problems encountered has been solved. Usually this requires more expensive design or operating techniques. Most often, problems in new systems are not encountered because the technology has failed, but because the proper equipment is more expensive than that installed.

4.1.4 Feed Preparation

The preparation of feed slurries has presented problems in lime, limestone, and magnesium oxide systems. Slaking, prior to fresh feed storage tanks, is necessary for the lime and magnesium oxide systems. Fine grinding has reduced problems in the use of limestone. The Japanese have had excellent results with limestone ground to less than 300-mesh. This has increased SO_2 removal and reduced plugging.

4.1.5 Solids Separation and Sludge Disposal

The disposal of sludge is a major consideration in selecting a nonregenerable FGD system. Land availability, sludge handling, and waste water contamination are potential problems with these systems. Increased SO_2 removal will tend to increase this problem since slightly more sludge will be generated. The impact on solid waste and environmentally acceptable disposal techniques is being discussed in detail in a separate study by the Aerospace Corp. for EPA.

Solutions to sludge disposal have been directed toward reducing its water content, reducing its volume, and increasing its compressive strength. A number of dewatering processes have been applied, and clarification, centrifugation, and vacuum filtration techniques predominate. In addition to these techniques, a number of plants are experimenting with or employing forced oxidation of sulfite to sulfate to enhance solids settling properties, to decrease sludge disposal land requirements, and to maximize the quality of recycled water. Two full-scale systems using forced oxidation are Northern States, Sherburne No. 1 and No. 2. Two additional full-scale systems using forced oxidation on a short-term experimental basis are Arizona Public Service, Cholla No. 1; and Commonwealth Edison, Will County No. 1. Systems planning to use forced oxidation for future operations are Northern States Power, Sherburne No. 3 and No. 4. Research is continuing in this area.

The two predominant methods of sludge disposal are onsite ponding and landfilling with physically conditioned or chemically treated waste. A number of specific disposal methods are available in each of these categories. The selection of a specific method depends upon a number of site-specific factors such as space, distance to site, etc. Currently, feasible alternates and trends include:

(a) Based upon systems now in service or planned for future operations, approximately 75 percent of the FGD operating capacity will be using on-site ponding for ultimate disposal of sludge. The remaining 25 percent capacity will employ physical and chemical treatment methods for landfill disposal.

(b) Two major suppliers now offer proprietary methods for chemically fixing sludge. The Dravo Corporation and IU Conversion Systems (IUCS) both market commercial stabilization processes. Dravo and IUCS supply systems that treat FGD waste from full-scale FGD systems. These systems are being used at the Phillips, Elrama and Conesville stations.

(c) A number of utilities operate FGD systems that include fly ash and lime stabilization of the thickener underflow, or vacuum filtration of solids prior to offsite landfill.

(d) A number of major FGD system suppliers now offer sludge treatment processes in addition to emission control systems. Research-Cottrell and Combustion Engineering, for example, now provide sludge treatment systems as separate packages from the FGD systems.

Water from the disposal pond is frequently recycled to the FGD system. Operation with no waste water disposal (closed loop) has proven difficult in some cases due to the need to purge dissolved solids from the system and the extra water needed at times to wash demisters. The entire water disposal problem is being investigated for EPA in a study by the Radian Corp.

4.2 SYSTEM OPERABILITY* AND AVAILABILITY**

4.2.1 Operability Trends

The design and operation of newer FGD systems benefits from the experience of older units. The observed trend of improved operability should therefore continue to increase. Based upon limited, utility-provided data, Figure 4-1 shows the cumulative FGD operability percentage as a function of

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* Hours FGD system operated/hours boiler operated.

** Hours FGD system was available/hours in that period.

FGD system start-up date. In this presentation, the individual unit availability or operability, from start-up to July 1977, is plotted as a function of plant start-up date. Operability is used in all cases except when all the flue gas must pass through the FGD unit (no by-pass). In that case availability is a more valid measure of operation. As will be noted, the trend line, which is a least squares regression fit, reflects greatly improved performance by newer FGD systems. Table 4-1 identifies the plants plotted in Figure 4-1.

Several examples of the improved availability and operability of FGD systems are noted below. Performance histories of these individual systems were presented in Section 3 and in the Appendices to this report.

Lime Systems

- Louisville Gas and Electric, Cane Run Unit No. 4, Louisville, Kentucky (178 MW)

  The unit began operation in August 1976. With the exception of a shutdown caused by the lack of lime (frozen rivers prevented barge deliveries), and process modifications, the unit has averaged in excess of 95 percent operability through July 1977.

- Louisville Gas and Electric, Paddy's Run Unit No. 6, Louisville, Kentucky (65 MW)

  The unit began operation in April 1973. Start-up problems, modifications, and an extended boiler shutdown kept operability low until October 1974. Operability from October 1974, to August 1977, has been in the 95- to 100-percent range, although the boiler operates primarily as a peak loading station.

- Kentucky Utilities, Green River Units No. 1, 2, & 3, Central City, Kentucky (64 MW)

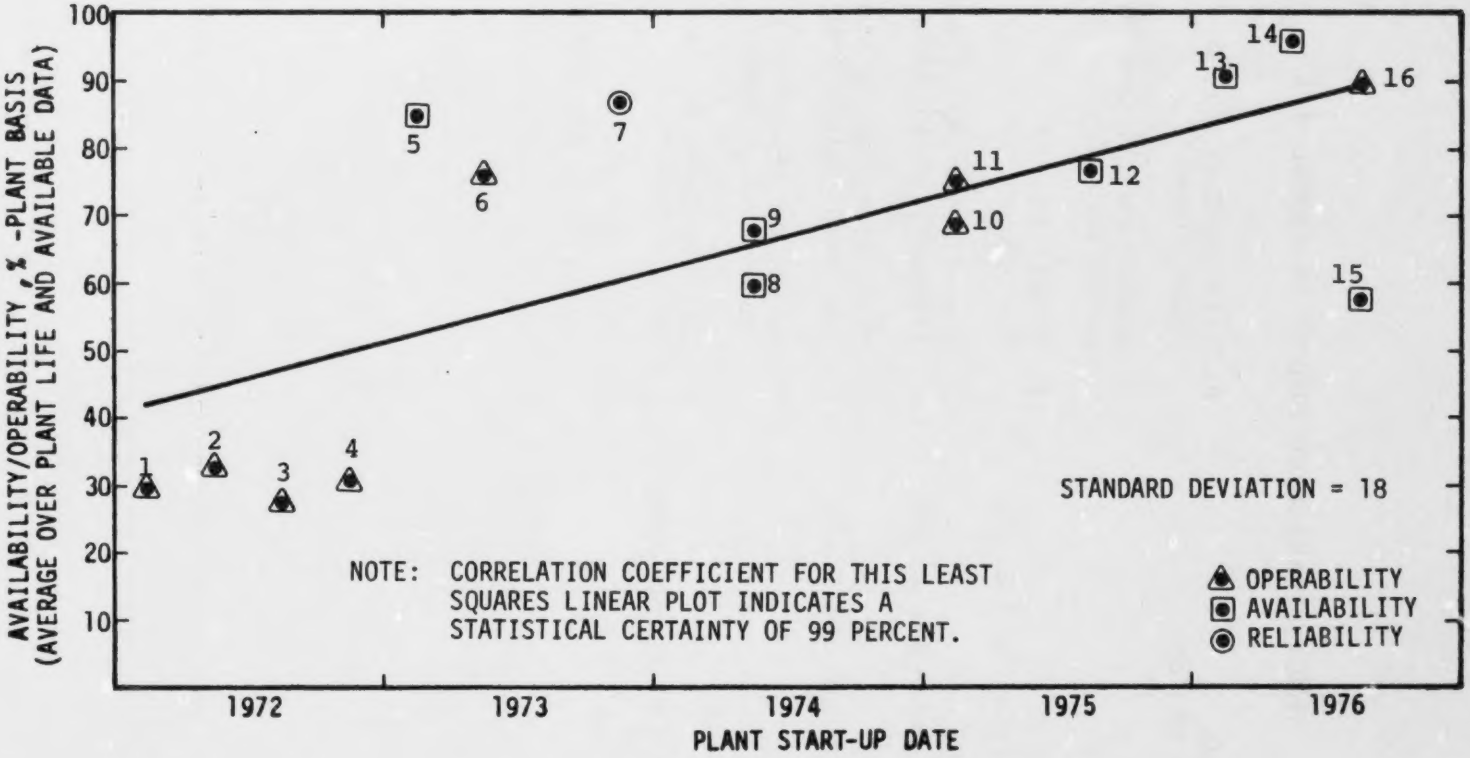

Figure 4-1. Average plant FGD availability/operability versus plant start-up date.

(See Table 4-1 for plant identification)

Table 4-1. IDENTIFICATION OF PLANTS IN FIGURE 4-1

1. Will County No. 1
   Commonwealth Edison
2. Mystic No. 6
   Boston Edison
3. Hawthorn No. 4
   Kansas City Power and Light
4. Hawthorn No. 3
   Kansas City Power and Light
5. LaCygne No. 1
   Kansas City Power and Light
6. Paddys Run No. 6
   Louisville Gas and Electric
7. Cholla No. 1
   Arizona Electric Power Co-Op
8. Reid Gardner No. 1
   Nevada Power
9. Reid Gardner No. 2
   Nevada Power
10. Scholz No. 1B and 2B
    Gulf Power Co.
11. Scholz No. 1A
    Gulf Power Co.
12. Green River No. 1 and 2
    Kentucky Utilities
13. Sherburne County Station No. 1
    Northern States Power Co.
14. Bruce Mansfield No. 1
    Pennsylvania Power
15. Reid Gardner No. 3
    Nevada Power
16. Cane Run No. 4
    Louisville Gas and Electric

This FGD unit began operation in December 1975. After the initial operating period (Dec. 1975-Feb. 1976), the operability of the unit was 97 percent through June 1977.

Limestone Systems

- Kansas City Power and Light, La Cygne Unit No. 1, La Cygne, Kansas (820 MW)

  Operation of this FGD system began in January 1974. Since all the flue gas must be treated by the FGD system, availability is a more meaningful parameter. From February 1976, through July 1977, unit availability averaged 93 percent.

- Northern States Power, Sherburne Nos. 1 & 2, Sherburne Minnesota (700 MW each)

  After start-up, No. 1 unit averaged 92-percent availability. No. 2 unit has averaged 95-percent since start-up.

Magnesium-Oxide Systems

- The first full-scale FGD unit (Philadelphia Electric, Eddystone No. 1) designed as a continuing operation was recently restarted after eighteen months of downtime for the forced relocation of the magnesium sulfite regeneration equipment. Two earlier installations (Boston Edison's Mystic Station and Potomac Electric's, Dickerson Station) were short-term, prototype units to explore the process potential and operating problem areas. Only data for the Mystic Station are shown in Figure 4-1.

In summary, the availability of full scale scrubbing facilities has increased steadily to where current systems are demonstrating long-term availabilities in excess of 90 percent, in most cases without the use of spare modules.

4.2.2 Availability

The ability of an FGD system to operate is as important as its efficiency in reducing SO_2 emissions. This ability to operate has been defined a number of ways including:

- Availability Factor - Hours the FGD system was available for operating (whether operated or not)/hours in period, expressed as a percentage.
- Reliability Factor - Hours the FGD system operated/hours FGD system was called upon to operate, expressed as a percentage.
- Operability Factor - Hours the FGD system was operated/boiler operating hours in period, expressed as a percentage.
- Utilization Factor - Hours FGD system operated/hours in period, expressed as a percentage.

Availability and operability are commonly interchanged in usage, and, if a boiler operates continually over a specific time period, they yield the same numerical value. Historically FGD system availability has varied depending on a variety of factors, and, naturally, individual module availabilities have varied even more so. Trends in overall system availability were shown earlier in Figure 4-1. These data demonstrated continuing and positive improvement in availability as newer units (with improved design and component features) came on-line. In addition, only one of these stations (Sherburne) has an extra or redundant module available for use in the event of an operating module malfunction or failure.

4.2.2.1 Module Availability - The range in average monthly module availabilities for a number of selected FGD systems, which historically have shown high average availability, is shown in Figure 4-2. The annual average availability for the overall FGD system is also shown. Because of the variations in system design, maintenance, and operational capability, it is difficult to generalize about module availability. It is, however, evident that a 90-percent average module availability can be achieved for both low- and high-sulfur coal applications.

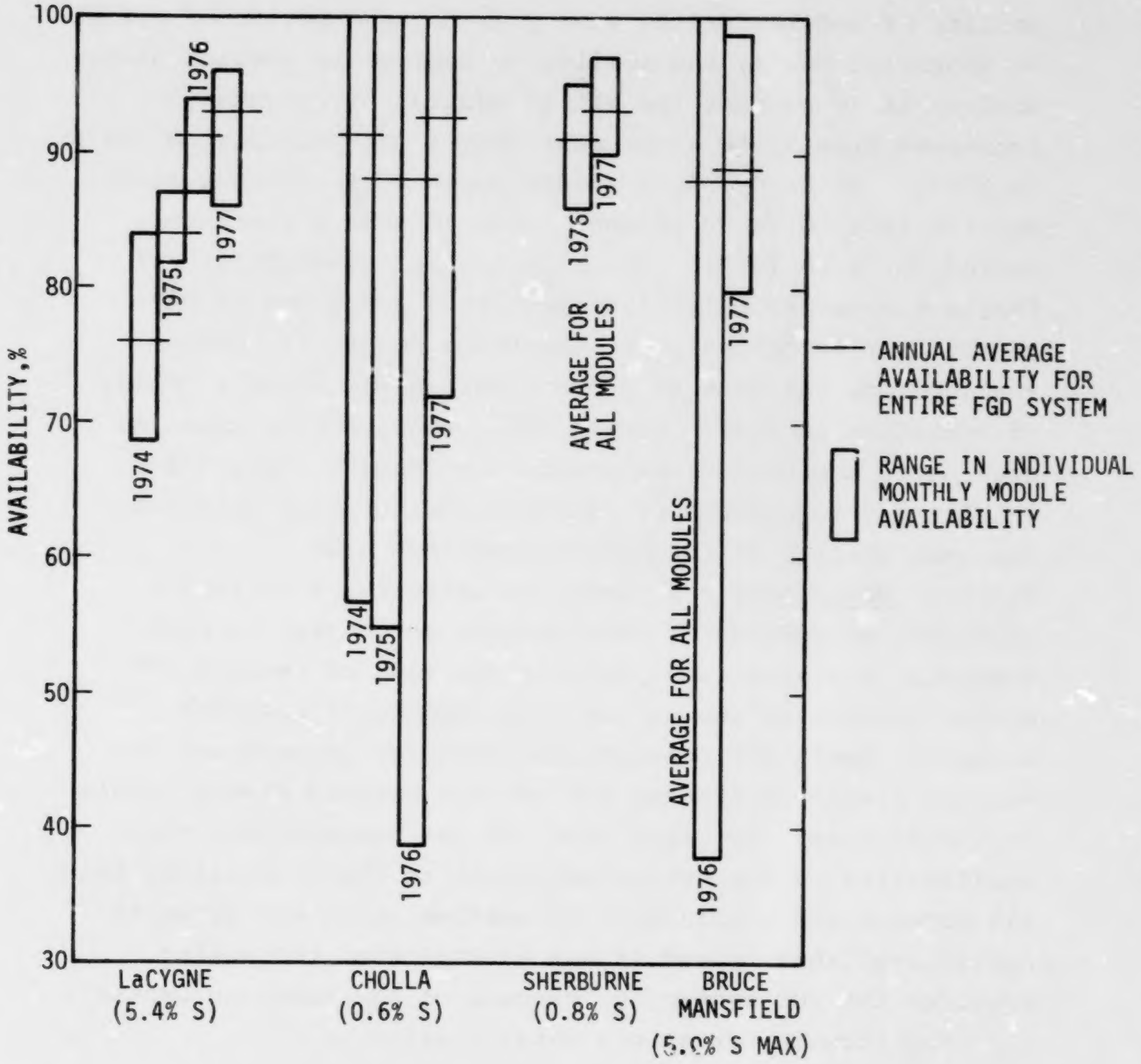

Figure 4-2. Average availability for selected FGD systems.

Also shown on this figure are trends in improved availability of modules at the same plant over a period of years. At Sherburne No. 1, the ability to achieve an average availability of 90 percent for all 12 modules (1 redundant) increased from 91 to 93 percent over a one-year period (1976 to 1977). At La Cygne, a steady increase in average availability from 76 to 93 percent occurred over a three-year period (1974 to 1977). The high initial availability of Cholla has varied only slightly over 4 years and is currently over 90 percent. At the Bruce Mansfield Station, availability was near 98 percent during the first 8 months of operation in 1976. During 1977, availability remained high until the boiler was shutdown in March. After restarting at 50 percent of capacity due to stack relining, the availability of the modules remained high.

4.2.2.2 Redundancy - In order to establish a basis for selection of redundancy requirements to be used in cost analysis, a statistical analysis was made of overall FGD system availabilities for various numbers of required modules. Table 4-2 presents availability projections for various sized FGD systems and various assumed single module availabilities. For this case, it was assumed that the availability of the system was equal to the probability that all modules are available. No consideration was given to boiler availability and it was assumed that the boiler required the FGD system 100 percent of the time (no bypass).

The formula for such a determination is

$$P_n = A^n,$$

Table 4-2. FGD SYSTEM AVAILABILITY PROJECTIONS BASED UPON NO SPARE MODULES

Boiler Size, MW	Minimum Number of Modules at Full Load	FGD System Availability Based Upon Single Module Availabilities of:					
		70	75	80	85	90	95
100 - 150	1	70	75	80	85	90	95
150 - 300	2	49	56	64	72	81	90
300 - 450	3	34	42	51	61	73	86
450 - 600	4	24	32	41	52	66	82
600 - 750	5	16	24	37	44	59	77
750 - 900	6	12	18	30	38	53	73
900 - 1050	7	8	13	24	32	48	70
1050 - 1200	8	6	10	19	27	43	66

where

P_n = probability that n modules are available,

n = number of modules, and

A = availability of a single module.

Inspection of the projected values indicates that a definite reduction in overall FGD system availability can be anticipated as the number of modules increases. To increase system availability, it is necessary to add spare (redundant) modules to compensate for those put out of service.

FGD system availabilities for various size systems with one or two spare modules are presented in Table 4-3. In the case of one spare module, the system availability is assumed to be equal to the probability that all modules are available (including the spare) plus the probability that the minimum required number of modules is available. This is given by the following formula:

$$P_{\geq n-1} = A^n + n(1-A)A^{(n-1)}$$

where

$P_{\geq n-1}$ = probability that (n-1) or greater modules are available.

This assumes that all modules have the same availability and are independent of each other.

For the case with two redundant modules, the system availability is assumed to be equal to the probability that all modules (including both spares) are available plus the probability that all modules but one are available plus the probability that the minimum number of required modules (two modules out of service) are available. This is given by the following equation:

Table 4-3. FGD SYSTEM AVAILABILITY PROJECTIONS BASED UPON USING ONE AND TWO SPARE MODULES (X + 1 and X + 2) AT FULL BOILER LOAD

Boiler Size MW	Minimum Number of Modules Required to Operate at Full Load (X)	System Availability Based Upon Single Module Availabilities of:											
		One Spare Module (X + 1)						Two Spare Modules (X + 2)					
		70	75	80	85	90	95	70	75	80	85	90	95
100 - 150	1	91	94	96	98	99	99+	97	98	99	99+	99+	99+
150 - 300	2	78	84	90	94	97	99	92	95	97	99	99+	99+
300 - 450	3	65	74	82	89	95	99	84	90	94	97	99	99+
450 - 600	4	53	63	74	84	92	98	74	83	90	95	98	99+
600 - 750	5	42	53	66	78	89	97	65	76	85	93	97	99+
750 - 900	6	33	44	58	72	85	96	55	68	80	89	96	99
900 - 1050	7	26	37	50	66	82	94	46	60	74	86	95	99
1050 - 1200	8	20	30	44	60	77	93	38	53	68	82	93	99

$$P_{\geq n-2} = A^{n} + n(1-A)A^{(n-1)} + \frac{n(n-1)}{2}(1-A)^{2}A^{(n-2)}$$

Based on this analysis, it is expected that if each module is available 90 percent of the time, one spare module will achieve an overall system availability of 90 percent up to a boiler size of about 750 MW. If each module is only 85 percent available, a single spare module will achieve system availability of 80 percent up to about 750 MW. With two spare modules, overall system availability is increased dramatically.

The previous analysis assumes that the particular FGD system is required 100 percent of the time. However, in practice coal-fired boilers have approximately an 80-percent requirement for availability.[3] This inherent down-time provides the opportunity for conducting preventative maintenance on the FGD system coincident with boiler servicing periods. This maintenance should increase the operability of the FGD system; that is, its capability to operate when the boiler is in operation. In addition to the above consideration, there is a reserve capacity margin for any public utility's entire system of generating plants and/or a regional power system of which one specific plant is a part. These reserve capacities further allow a particular boiler to reduce load and service its FGD module(s) without interrupting the power supply. Typical percent reserve margins for 1977 range from 18.7 to 50.9 percent during summer peak periods and from 21.7 to 93.6 for winter peak periods.[4] Total U. S. values are 29.2 percent for summer peak demand and 46.9 percent for winter peak demand.

Additionally, during times of forced outage on an FGD system, there usually are "peaking" units within the particular utility system which can be used to provide the makeup power required to cover the resultant partial load operation. In case there are no other generating units available within a particular utility system to meet the outage demand, power can be purchased from another company within the "grid" system which provides a regional back-up reserve power margin.

It can be readily seen that a valid analysis of FGD system redundancy needs to take many factors on system reliability into consideration.[5,6,7] Such an analysis would require specific data for the utility company under study and its interconnection with other companies in the same grid. This analysis cannot be done within the scope of this effort, however an analytical approach is outlined as follows:

> Assume a reserve margin of 29 percent based on the results reported by FPC.[6] Furthermore, assume that the plant size is 1000 MW, and eight scrubber modules are installed (one spare). With one scrubber module down, the plant can operate at full load. If two modules are down, the plant can serve 6/7 or 85.7 percent of the full load, and if three are down, 5/7 or 71.4 percent of the full load. The probabilities of these different situations, assuming an individual module availability of 90 percent, are given in Table 4-4.
>
> Thus, this particular FGD system will provide full-load service (no outage) at least 81 (43+38) percent of the time and 85.7 (6/7) percent of full load for an additional 15 percent of the time. Assuming the 29 percent reserve margin, this FGD system can provide more than 85 percent of the load for 96 (81+15) percent of the time. The FGD system will serve at least 71.4 percent of full load 99.5 percent of the time.

Conclusion: Thus, there would be a deficiency, with respect to serving at least 71 percent of the full load, only 0.5 percent of the time.

Assuming a module availability of 80 percent, the probabilites shown in Table 4-5 for corresponding outages are obtained.

In this case, the FGD system would serve at least 71 (5/7) percent of the full load for 94.5 percent of the time and would be deficient 5.5 percent of the time.

Conclusion: One spare module would be considered adequate under the assumption that each module has 80-percent availability and that there is a 29-percent reserve margin.

4.2.2.3 Manufacturer's Availability Guarantees - Seven out of twelve FGD system suppliers indicated a willingness to guarantee overall availability of their systems. Ninety percent was the typical "level-of-performance" guaranteed as shown in Table 4-6. In most cases the availability is negotiable and subject to guarantees from component suppliers for varying amounts of time up to two years.

In addition to SO_2 removal efficiencies and availabilities, vendors are guaranteeing additional parameters. For instance, the following items are included in the guarantee at the Conesville Power Plant in Ohio:

- lime usage rate, lbs/hr
- makeup water rate, lb/hr
- pressure drop, inches of water
- particulate emissions, no increase due to scrubber
- system availability during a 6-month demonstration test, 90 percent overall.

4.2.3 Summary

The ability to operate at a high level of availability has been demonstrated by a number of FGD systems on both high- and low-sulfur coal. High availability is largely a function of design and maintenance, and as more operating

Table 4-4. PROBABILITIES AND OUTAGES FOR SPECIFIED NUMBER OF SCRUBBER MODULES OPERATING, ASSUMING MODULE AVAILABILITY OF 90 PERCENT - 1000-MW BOILER

Number of scrubber modules available	Corresponding probability	Corresponding outage, MW
8	$(0.90)^8 = 0.430 = 43\%$	0
7	$8(0.90)^7(0.10) = 0.383 = 38\%$	0
6	$28(0.90)^6(0.10)^2 = 0.149 = 15\%$	143
5	$56(0.90)^5(0.10)^3 = 0.033 = 3.3\%$	286
< 5	0.005	≥ 286
Total Probability	1.000	

Table 4-5. PROBABILITIES AND OUTAGES FOR SPECIFIED NUMBER OF SCRUBBER MODULES OPERATING, ASSUMING MODULE AVAILABILITY of 80 PERCENT - 1000-MW BOILER

Number of scrubber modules available	Corresponding probability	Corresponding outage, MW
8	$(0.80)^8 = 0.168 = 17\%$	0
7	$8(0.80)^7(0.20) = 0.336 = 34\%$	0
6	$28(0.80)^6(0.20)^2 = 0.294 = 29\%$	143
5	$56(0.80)^5(0.20)^3 = 0.147 = 15\%$	286
≤ 4	$= 0.055 = 5.5\%$	≥ 429
Total Probability	1.000	

Table 4-6. SUMMARY OF AVAILABILITY GUARANTEES OFFERED BY MANUFACTURERS[a]

Company[b]	Guarantee offered	
	Yes (level)	No
A	Normally better than 90%	
B		X
C	Typically 90% during performance testing; sometimes up to 95%	
D	Maximum of 90% based on boiler hours	
E	Yes (level of guarantee not disclosed)	
F	Have guaranteed in excess of 90%	
G	Normally 85 to 90% for 1 or 2 years	
H		X
I		X
J	Maximum of 90% on a case-by-case basis	
K		X
L		X

[a] Represents the responses of 12 manufacturers.

[b] Company names are deliberately withheld.

experience is gained, the methods for achieving higher availabilities are being implemented. Installations of an extra or redundant scrubber module can significantly increase overall system availability since repairs can be made with the boiler in operation.

4.3 SYSTEM EFFICIENCY

As presented in Chapter 3, all five of the demonstrated FGD systems (lime, limestone, double alkali, magnesium oxide, and Wellman-Lord) have achieved SO_2 removal efficiencies in excess of 90 percent. Systems which have reported 90 percent SO_2 removal or greater are identified in Table 4-7 by utility, station, unit no., removal efficiency, and nature of application.

In Japanese FGD installations, these same processes are reported to have demonstrated SO_2 removal efficiencies well in excess of 90 percent, many in excess of 95 percent. However, only a few of these are on coal fired boilers.

Extensive pilot, prototype, and demonstration scale data have been reported to document the complex relationships between process parameters and SO_2 removal efficiency. Among the test facilities are:

Shawnee	TVA	10 MW	Lime/Limestone
Mohave	So. Calif. Edison	170 MW	Lime/Limestone
Four Corners	Arizona Public Ser.	160 MW	Lime
Mystic	Boston Edison	150 MW	Mag. Oxide
St. Clair	Detroit Edison	163 MW	Limestone
Scholz	Gulf Power	20 MW	Double alkali
		20 MW	Activated carbon

The overall removal efficiency capability has progressed to the point that almost all vendors will guarantee SO_2 removal efficiences in excess of 90 percent, as shown in Table 4-8.

Table 4-7. PLANTS REPORTING 90 PERCENT OR GREATER SO_2 REMOVAL

Utility Company	Station	Unit	Nature	% SO_2 Removal
Arizona Public Service	Cholla	No. 1	Full-scale	92
	Four Corners	No. 5	Demonstration	95
Duquesne Light	Phillips	Nos. 1-6	Full-scale	90+
Louisville Gas & Electric	Cane Run	No. 4	Full-scale	90
	Paddy's Run	No. 6	Demonstration	99.5
Northern Indiana Public Service	Mitchell	No. 11	Demonstration	90
Philadelphia Electric	Eddystone	No. 1	Demonstration	95-98
Tennessee Valley Authority	Shawnee	No. 10	Prototype	95-99
Boston Edison	Mystic	No. 6	Demonstration	90
Detroit Edison	St. Clair	No. 6	Demonstration	90-91
General Motors	Parma	No. 1-4	Full-scale*	90+
Gulf Power	Scholz	No. 1	Prototype	95
	Scholz	No. 1-2	Prototype	95
	Scholz	No. 2	Prototype	90+
Potomac Electric & Power	Dickerson	No. 3	Demonstration	90
Southern California Edison	Mohave	No. 1	Demonstration	95
	Mohave	No. 2	Demonstration	95
U.S. Air Force	Rickenbacker	No. 1-9	Full-scale**	99
Kentucky Utilities	Green River	No. 1-3	Full-scale	90+

* Industrial

** Military Base

Table 4-8. GUARANTEES OFFERED BY MANUFACTURERS FOR SO_2 REMOVAL

Company[a]	Level of SO_2 removal guaranteed		
	<90	90%	>90%
A		Minimum guarantee given	Is willing to offer 95% guarantee on case by case basis
B		Minimum guarantee given	For >90%, it is based on inlet SO_2 concentration
C			Would guarantee 95% in all cases
D			Would guarantee up to 92%
E			Have guaranteed >90% in the past
F		This guarantee is normally given	Depending upon the process, they would guarantee >90%
G		This guarantee is given where SO_2 inlet concentration is 500-4,000 ppm	Have guaranteed up to 95% in the past
H	Would normally guarantee 80-85%	This guarantee is given where low sulfur coal is utilized	
I		Minimum guarantee given	Are prepared to offer better than 90% with low or high sulfur coal, but would not guarantee less than 50 ppm SO_2 concentration in exit stream
J		This guarantee is usually given with coal having 3-4% sulfur	
K		This guarantee is normally given with low or high sulfur coal	In many cases they guarantee 95% with high sulfur coal
L		Minimum guarantee given	May guarantee up to 95% on a case by case basis

[a] Company names are deliberately withheld.

4.3.1 Lime and Limestone Systems

Many of the current operating lime and limestone scrubbing systems were designed for SO_2 collection efficiencies under 90 percent to satisfy an applicable regulation. Often, an efficiency in the range of 60 to 70 percent was sufficient to meet the specific emission limitations and therefore, values of this magnitude were used for design criteria. Design of newer systems which are required to achieve high efficiency must take into account a number of key design variables including:

- inlet SO_2 concentration rate
- liquid to gas ratio
- scrubber gas velocity
- scrubber liquor inlet pH
- type of absorber
- magnesium content
- type of alkali

Inlet SO_2 concentration and rate - Higher removal efficiencies can be more easily achieved at lower SO_2 inlet concentrations because the amount of SO_2 that must be absorbed per unit of scrubbing liquor to achieve a specified outlet concentration is smaller. When the scrubbing liquor contacts a high SO_2 concentration it quickly reacts with the SO_2, decreasing further absorption capability. Since the dissolution rate of the alkali in the scrubber media is slower than the reaction rate with SO_2, a "back pressure" or resistance to further absorption is created which can inhibit a scrubber efficiency. At low SO_2 concentrations the alkali in the liquor can react with a greater percentage of the SO_2 and affect a greater removal efficiency under a given set of operating conditions.[8,9]

As the back pressure of SO_2 in solution approaches the partial pressure of SO_2 in the flue gas, absorption decreases rapidly. However, by contacting the lowest SO_2

bearing gas with the fresh slurry (countercurrent flow), high efficiency can be maintained. At the Mohave Station greater than 95 percent efficiency was achieved at an SO_2 inlet concentration of 200 ppm with both lime and limestone and with three different types of absorbers. The TVA Shawnee facility reports 95 percent removal at high SO_2 inlet concentrations.

Under typical plant operation, the inlet SO_2 concentrations will vary directly with changes in fuel sulfur content. To maintain a desired level of efficiency, variable amounts of alkali must be brought into contact with the flue gas. This can be accomplished rapidly by directly varying the liquor flow rate based on the inlet SO_2 readings, which is a "feed-forward" type of control system. That is, the scrubbing system responds to a signal which indicates SO_2 concentrations before the scrubber.

Liquid to gas ratio - Higher efficiencies are realized at higher L/G ratios for lime and limestone systems, as previously shown in Figures 3-11 to 3-13, and 3-30 and 3-31, respectively. For a given absorber, increased L/G ratios will yield higher efficiencies until flooding and poor gas distribution occur. For new designs, absorbers which can accomodate high L/G ratios can be selected and high efficiency maintained. Higher L/G ratios also require larger pumps, pipes, and slurry reaction tanks. Again, these can be designed into the system and should cause no unusual operating problems. Additionally, scaling potential will actually decrease since the slurry contains less SO_2 per unit volume.

The following L/G ratios have yielded efficiencies in excess of 90 percent:

Spray Tower: Lime or limestone - greater than 10. l/m^3 (80 gal/1000 acf)

Turbulent Contact Absorbers: Lime - 8 l/m^3 (60 gal/1000 acf)

Turbulent Contact Absorbers: Limestone - 10.7 l/m^3 (80 gal/1000 acf)

Scrubber Gas Velocity - The effects of changes in flue gas absorber velocity on SO_2 removal efficiency, when other variables are kept constant, vary with the type of absorber. For a spray tower, the efficiency decreases at a fixed L/G ratio (Figure 3-21). This effect is much less noticeable on packed and turbulent contact type absorbers where increased agitation and mixing partially off-set the decreased L/G ratio. Velocities in the range of 2.1 to 2.4 m/s (7 to 8 ft/s) and 2.7 to 3 m/s (9 to 10 ft/s) appear desirable for spray tower and turbulent contact absorbers, respectively.

Scrubber liquor inlet pH - This parameter has a direct and noticeable effect on SO_2 collection efficiency as shown previously in Figures 3-17 and 3-32 for lime and limestone slurries, respectively. Increased efficiency is achieved at higher pH since more alkali is available and higher dissolution rates are achieved (as previously noted, operation at very high pH, however, causes scaling problems).

Based on tests at EPA's Shawnee Test Facility, a pH of 5.8 and 8.0 should be used for limestone and lime slurry systems, respectively. Operation at these pH levels requires lime and limestone in excess of that required to chemically react with the SO_2 (about 20 percent excess for lime and 40 percent excess for limestone). Maintenance of the desired pH by careful measurement and close control of reagent feed and mixing system will prevent the pH variations which reduce efficiency (if too low) or cause scaling (if too high).

Type of absorber - A large variety of absorber designs have been utilized to achieve SO_2 removal efficiencies as high as 99 percent. These include cross-flow horizontal spray chambers (Weir), spray towers, packed-grid towers, and TCA (mobile bed) absorbers. The venturi type has also been used, however, it is more useful as a particulate removal scrubber and not as efficient for SO_2 absorption. The final selection and design of an absorber is usually based on previous test data and on the required liquid and gas flow rates. Spray towers (either horizontal or vertical) offer a number of advantages including simple internal design which decreases scaling potential, acceptance of high liquid flows and decreased maintenance as reported at the Mohave and Shawnee plants.

Magnesium Content - The addition of relatively small amounts of soluble magnesium (less than 1 percent by weight) to the scrubber liquor in the form of magnesium oxide, magnesium sulfate, or dolomitic lime (in lime systems) can greatly increase the SO_2 collection efficiency of the system. Magnesium compounds are much more soluble, compared to calcium, and can react rapidly in the liquid phase with SO_2. The effect of magnesium addition was shown in Figures 3-18 and 3-34 for lime and limestone systems respectively, and was tested at the Paddy's Run plant and Phillips plant (lime based FGD systems). These data show SO_2 collection efficiencies in excess of 90 percent when the effective magnesium concentration is over 0.5 percent. Any chloride present in the liquor will react with the magnesium to form inactive magnesium chloride. Additional magnesium must be added, above that which will react with any chlorides, to yield an effective or available magnesium concentration. Since soluble magnesium compounds can be lost in the sludge, possible treatment of any water discharge and addition of fresh magnesium to replenish losses is required.

Two proprietary absorbents have been developed, one by Dravo (Thiosorbic lime: 2 to 6 percent magnesium oxide lime), the other by Pullman Kellogg (Catalytic limestone: 3 to 27 percent magnesium sulfate limestone). Three full-scale, operational FGD systems (Columbus and Southern Ohio Electric, Conesville No. 5; Pennsylvania Power, Bruce Mansfield No. 1 and No. 2) are now using Dravo's thiosorbic lime reagent. In addition, six more full-scale systems plan to use thiosorbic lime in future FGD operations:

Big Rivers Rural Electric Power Cooperative, Reid No. 2
Columbus and Southern Ohio Electric, Conesville No. 6
Duquesne Light, Elrama Nos. 1-4
Duquesne Light and Phillips Nos. 1-6
Indianapolis Power and Light, Petersburg No. 3
Pennsylvania Power, Bruce Mansfield No. 3

Type of alkali - Both lime and limestone are effective compounds for SO_2 removal. Lime is more reactive and lime scrubbing systems are more efficient than limestone systems under a given set of conditions. However, since the type of alkali must be initially considered in the design phase, either system can be engineered to yield high efficiencies.

A model has been developed by Bechtel, based on their tests at the Shawnee pilot plant, that incorporates the various factors that affect lime and limestone scrubbing systems. These models are useful for selecting design factors for conditions within the model's bounds.[10]

4.3.2 Sodium Carbonate and Double Alkali FGD Systems

There are currently three sodium carbonate scrubbing systems in operation. These systems do not recycle any sodium carbonate but operate on a "once-through" basis (see Section 3.6.7). Double alkali systems, which produce a sludge and recycle the absorbent solution, are in operation in a number of large industrial sized boilers, but not on utility sized units (Section 3.3). Because of the more reactive nature of sodium carbonate and its solubility in water, the sodium carbonate scrubbing systems have in general experienced high efficiencies (greater than 90 percent at Reid Gardner) and few operating problems in the absorber area.

Based on the limited application of these systems, the following guidelines should be followed to achieve high efficiency:

(a) Utilization of a prescrubber with a separate water recirculating system for control of particulates and chlorides for high chloride coal (>0.04 percent Cl by weight in the coal).

(b) The use of a two-stage tray or packed tower absorber with an L/G of 1.3 to 2.7 l/m^3 (10 to 20 gal/1000 acf). Typically the absorber pressure drop is 15 to 30 cm (6 to 12 inches) of water.

(c) The absorber scrubbing liquor pH being recycled to the absorber should be in the 6.0 to 7.0 pH range.

(d) If lime regeneration is used in a double alkali system, the reaction tank residence time should be approximately 10 minutes. Lime utilization of about 90 percent may be expected.

(e) If limestone regeneration is used, the reactor tank residence time should be approximately 30 minutes. Limestone utilization of 75 to 85 percent may be expected.

(f) Carbonate softening and sodium ion make-up can be accomplished through addition of sodium carbonate (soda ash).

4.3.3 Magnesium Oxide FGD Systems

This system has received considerable attention in pilot plant studies and is currently installed on one coal fired power plant in this country and two industrial applications in Japan (Section 3.4). Due to the reactive nature of magnesium oxide, efficiencies in excess of 90 percent have been achieved with various types of absorbers. Based on experience at the Mystic and Dickerson Stations, the following conditions should be used as guidelines to achieve high efficiency:

(a) High efficiencies (99%) particulate removal should be accomplished by use of ESP(s) or venturi scrubber(s).

(b) A prescrubber should be used to remove any remaining particulate and most of the chlorides and SO_3.

(c) Utilize venturi absorbers, typically operating at a pressure drop of 25 cm (10 inches) of water or greater, or Turbulent Contact Absorbers operating at approximately 20 cm (8 inches) of water pressure drop, at an L/G of 5.3 to 6.7 l/m^3 (40 to 50 gal/1000 acf).

(d) The absorber superficial gas velocity should not exceed approximately 3.0 m/sec (10 ft/sec) range.

(e) The slurry pH measured at the absorber discharge should be maintained in the 6.0 to 7.5 range.

4.3.4 Wellman-Lord FGD Systems

Experimental data and full scale operating data on the single Wellman-Lord FGD system on a coal-fired boiler in this country have demonstrated levels of SO_2 collection efficiency in excess of 90 percent with 3.5 percent-sulfur coal. The following conditions should be used as guidelines to obtain high efficiency.

(a) Install a prescrubber with a separate water recirculation system for final particulate control and reduction of SO_3 and chlorides.

(b) Use a three to five tray absorber with an L/G of 1.0 to 1.3 l/m^3 (6 to 10 gal/1000 acf). Typically the absorber pressure drop is 27 to 44 cm (10.5 to 17.5 inches) of water.

(c) A superficial gas velocity in the range of 2.7 to 3.1 m/sec (9 to 10 ft/sec) should be used.

(d) Maintain the recycled sodium sulfite scrubbing solution at a pH of 6.0 at the absorber inlet.

(e) The system makeup of fresh, 20-percent sodium carbonate solution should be approximately 0.07 l/m^3 (0.5 gallon/1000 acf) per tray.

(f) As the inlet SO_2 concentration decreases, the number of trays required to obtain high SO_2 removal should be increased.

4.3.5 Summary

The ability to achieve 90 percent SO_2 removal efficiency has been demonstrated by lime, limestone, double alkali, Wellman Lord and magnesium oxide FGD systems. However, all FGD systems have not demonstrated equal capability to achieve high efficiency on both high- and low-sulfur coal. Thus, for a particular application proper scrubber selection and design is required to achieve high efficiency.

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APPENDIX A

DOMESTIC LIME SLURRY FGD SCRUBBING SYSTEMS

This appendix summarizes data on the three lime scrubbing FGD systems presented in Section 3.1 and also describes the other operating lime systems in this country.

The following FGD systems are described:

- Bruce Mansfield No. 1
- Louisville Gas and Electric - Cane Run No. 4
- Montana Power Co. - Colstrip Nos. 1 and 2
- Columbus and Southern Ohio Electric Co. Conesville No. 5
- Duquesne Light Co. - Elrama Power Station
- Arizona Public Service - Four Corners No. 5
- Kentucky Utilities - Green River Boilers 1, 2, and 3
- Southern California Edison - Mohave
- Louisville Gas and Electric Co. - Paddy's Run No. 6
- Duquesne Light Co. - Phillips Power Station No. 6
- Rickenbacker Air Force Base
- Tennessee Valley Authority - Shawnee No. 10A and 10B

PENNSYLVANIA POWER
BRUCE MANSFIELD NO. 1

BACKGROUND INFORMATION

The Bruce Mansfield plant is a 2700-MW, three-boiler, coal-fired facility located on the Ohio River in Shippingport, Pennsylvania. This facility was built by Pennsylvania Power Co., which is acting on its own behalf and as an agent for the other participating companies, the Cleveland Electric Illuminating Co., Duquesne Light Co., Ohio Edison Co., and Toledo Edison Co.

Bruce Mansfield Nos. 1 and 2 are coal-fired, once-through, supercritical steam generators, which fire 333 tons/hr of coal and generate approximately 6.5 MM lb/hr (each) of steam at 3785 psig and 1005°F. The emission control equipment required for this unit is designed to meet state emission regulations of 0.6 lb sulfur dioxide/10^6 Btu of heat input and 0.0175 gr/scf of particulate when burning 11,900 Btu/lb coal having average ash and sulfur contents of 12.5 and 4.7 percent, respectively. Additional design-related information is presented in Table A-1.

Table A-1. FGD SYSTEM DATA, BRUCE MANSFIELD NO. 1

Unit location	Shippingport, Pennsylvania
Unit rating	825 MW
Fuel characteristics	Coal with 4.7 percent sulfur content (average)
FGD vendor	Chemico
Process	Lime scrubbing
New or retrofit	New
Start-up date	4/76
Efficiency, particulates	99.8 percent (design)
Efficiency, SO_2	92 percent (design)
Sludge disposal	Stabilized sludge is disposed of in unlined reservoir

POLLUTION CONTROL

The emission control system is a venturi wet-scrubber system for removal of fly ash and sulfur dioxide. It was manufactured by Chemico Air Pollution Control Co. and utilizes Dravo Corporation's Thiosorbic lime[a] as the scrubbing absorbent. The emission control systems for Boilers 1 and 2 are identical. The system planned for Boiler 3 in 1978 will utilize a Weir Horizontal Scrubber system.

Each scrubbing train consists of a variable-throat venturi, a 9000-hp I.D. fan, and a fixed-throat venturi in series. There are six scrubber trains per boiler. The design called for five trains to handle the flue gas flow from the boiler at full load, but due to higher flue gas flows, all six are required. The scrubbed flue gas is reheated by a direct oil-fired burner and then exhausted through a 950-ft stack.

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[a] Dravo's patented lime which contains 6 to 12 percent magnesium oxide.

The variable-throat venturi removes fly ash contained in the flue gas and the absorber module removes the remaining fly ash. Sulfur dioxide is absorbed in both the particulate scrubber and the absorber by the lime slurry, which contains 2 to 6 percent magnesium oxide. The unit is guaranteed to remove 92 percent of the sulfur dioxide and 99.8 percent of the particulate matter from the flue gas.

The scrubber discharge stream is combined with fly ash slurry and discharged to a 200-foot-diameter thickener. Sludge from the thickener underflow is pumped to a waste disposal system, where it is mixed with a stabilizing agent (Calcilox), and then pumped approximately 7 miles to the Little Blue Run Ravine for landfill.

The initial shakedown and debugging phase of operation began for part of the system in December 1975. Full commercial operation commenced in May 1976.[1]

PERFORMANCE HISTORY

This facility reported 100 percent operability during the first seven months after start-up. Operating problems were solved without causing boiler downtime. However, during the cold winter months of January and February 1977, the boiler lost 11 and 24 percent generating capability, respectively, because of extreme weather conditions. On March 21, 1977, the boiler was shut down for a 10-week turbine overhaul. During this time, work began on replacing the polyester flakeglass lining in the exit flue serving three scrubber modules. Failure resulted from poor application of the original lining, which was accomplished by spraying, and probably from the lack of flue gas reheat and poor mist elimination. The new lining is being troweled on. The other chimney flue, for the D, E, and F modules, also requires repair. Roughly 1 year will be required to complete this work, and the boiler will be held to half load for that time. Also, during the reduced load operation, Pennsylvania Power will be making other system changes as described in the following paragraphs.[2,3]

Mist Eliminator - The company has experienced excessive mist carry-over during scrubber operation. The mist eliminator was designed for 1 gr/scf liquid carry-over but an actual carry-over of about 3 gr/scf has been estimated by plant personnel. It is also designed to operate at gas velocities of 8 to 10 ft/sec. Pennsylvania Power would like to install a vertical mist eliminator in the ducting downstream of the scrubber vessels. Because of duct diameter and space restrictions, however, such a design would experience flow velocities as high as 50 ft/sec. A vertical mist eliminator was installed but collapsed as a result of structural failure caused by high flue gas velocity. The problem of plugging was also apparent, but it has been solved by relocating the lime slurry feed to the scrubber to a lower position.

Closed Loop - The system is not being operated closed-loop, because water is being retained at the sludge disposal site and not recycled to the process. Makeup water from the disposal pond is not needed since "fresh" water is added to the system in the fan sprays and during lime slaking. The company thinks they can operate closed loop, but have not as yet because they have been concentrating on the system's mechanical problems. Also, in the event they do not operate closed loop, the plant has a permit to discharge water to the Ohio river. Discharge requirements for water are total suspended solids of 60 mg/l or less; chloride content of 10 mg/l or less; and a pH between 6 and 9.

I.D. Fans - Pieces of scale carried through the system have caused the rubber lining of the fan scrolls to be chipped away, resulting in exposure and corrosion of the carbon steel. These fan scrolls are now being replaced with ones manufactured of Inconel. Pitting has also occurred to the carbon steel hubs because of seal leakage. The fan shafts are fabricated of carbon steel clad with Carpenter 20.

Reheater - The reheaters are not used because of duct vibrations caused by resonance. The reheaters are now undergoing design changes to correct the resonance problem. Plume buoyancy appears to be sufficient without reheat, but on certain days atmospheric conditions cause condensation and precipitation of moisture from the plume. Winds are such that the liquid fallout occurs in Shippingport. The fallout occurs as a clear liquid, but leaves a film upon drying. The pH of the condensed liquid is about neutral.

Scrubber - Scaling of the system has not been a serious problem. Scale has built up to about 1/2 to 1 inch over a 1-year period. Some plugging of the venturi nozzles with scale has occurred. This problem was eliminated by installation of "baskets" in the main header to catch the scale particles upstream of the nozzles. The baskets can be cleaned without shutting the system down. Maintenance of the scrubber system is not performed on a preventive basis but is done on the basis of need. Scheduled maintenance occurs every 12 months.

Performance Test - The Pennsylvania Department of Environmental Resources tested the unit on July 19 and 20, 1977.

SO_2 emissions were 0.44 lb $SO_2/10^6$ Btu and 1.26 $lb/10^6$ Btu, respectively. This represents 94 and 83 percent removal. The allowable emission rate is 0.6 $lb/10^6$ Btu. Since these are recent tests, only the results are available at this time. The details and causes for such a wide variation in results have not been published.[24] EPA, OAQPS, is obtaining test data from this facility by using continuous monitors. These data will be presented in a separate EPA report.

LOUISVILLE GAS AND ELECTRIC
CANE RUN NO. 4

BACKGROUND INFORMATION

The Cane Run Power Station, located in Louisville, Kentucky, is operated by the Louisville Gas and Electric Company. The plant has six electric power steam generating units, providing a total steam turbine net generating capacity of 992 MW.

Unit No. 4 is a coal-fired boiler with a continuous net generating capacity of 178 MW. The unit has a maximum power generation capacity of 190 MW. The unit heat rate is 10,030 Btu/kWh. The boiler is currently burning coal with a gross heating value of 11,500 Btu/lb and average sulfur and ash contents of 3.5 to 4.0 percent and 11 to 12 percent, respectively.

POLLUTION CONTROL

The emission control system for this unit consists of an electrostatic precipitator (ESP) upstream of a wet scrubbing system. The ESP provides primary particulate control while the wet scrubbing system provides additional particulate removal and primary sulfur dioxide control.

The FGD system consists of two identical parallel scrubbing trains designed and installed by the American Air Filter (AAF) Company. The wet scrubbing system utilizes a slurry of carbide lime (a waste by-product obtained from a nearby acetylene manufacturing plant). The hydrated lime contains 90 to 92 percent calcium hydroxide, 2.0 to 2.5 percent silica, 3 to 8 percent calcium carbonate, and 0.1 percent magnesium oxide. Design data are presented in Table A-2.

Table A-2. FGD SYSTEM DATA, CANE RUN NO. 4

Unit location	Louisville, Kentucky
Unit rating	178 MW
Fuel characteristics	Coal 3.5 to 4.0 percent sulfur
FGD vendor	American Air Filter
Process	Lime scrubbing
New or retrofit	Retrofit
Start-up date	August 1976
Efficiency, particulates	99 percent (guaranteed)
Efficiency, SO_2	85 percent (guaranteed)
Sludge disposal	Stabilized sludge is disposed of in unlined pond

Each scrubbing train is equipped with a guillotine-type bypass damper allowing bypassing of the gas around the scrubbers. Each scrubbing train contains the following major pieces of equipment: I.D. booster fan, quench section, flooded elbow, mobile bed contactor, centrifugal demister, and a three-section reactant tank system. The waste disposal system consists of a 75-foot-diameter thickener for liquid-solids separation and a fly ash pond for ultimate disposal of the thickener underflow. The scrubbing wastes are stabilized with fly ash, and water is recycled back to the process. The system does not use stack gas reheat. A wet stack lined with Carbolyne obviates the necessity of reheat.[4]

PERFORMANCE HISTORY

Available data on operability of the FGD system at Cane Run No. 4 are given in Figure 3-24. These data show a high level (90%) of operation during the first four months of operation, before the lime supply was curtailed. Operation in early 1977 was very limited before the unit was shut down for modification; which consisted of the addition of turning vanes in the absorber and conversion to a chevron-type mist eliminator.

There are no published SO_2 efficiency data. In response to a questionnaire from the EPA in June of 1977, the utility reported preliminary SO_2 removal efficiencies of 80 percent.[5] These data were taken before the absorber gas distribution modifications were made. The EPA is currently running more definitive efficiency tests, which will be documented in a separate EPA report.

MONTANA POWER CO.
COLSTRIP NOS. 1 AND 2

BACKGROUND

The Colstrip Power Station located near Colstrip, Montana, is operated by the Montana Power Company. The station consists of two, relatively new pulverized-coal-fired boilers. Each boiler is rated at 360 MW of power generating capacity. The station is located at the mouth of a mine and burns low-sulfur coal. Further details can be found in Table A-3. Because of the relatively low sulfur content of the coal (0.8%), a design efficiency of only 60 percent was utilized.

POLLUTION CONTROL

Both Unit No. 1 and Unit No. 2 are equipped with FGD systems. The systems are Arthur D. Little/Combustion Equipment Associates (ADL/CEA) lime/alkaline fly ash scrubbing systems designed to remove both particulate matter and sulfur dioxide from the flue gas. Design removal efficiencies are 99 percent for particulate matter and 60 percent for sulfur dioxide, as shown in Table A-3. The efficiencies are sufficient to meet existing applicable codes.

Table A-3. FGD SYSTEM DATA, COLSTRIP NOS. 1 AND 2

Unit location	Colstrip, Montana
Unit rating	360 MW each
Fuel characteristics	Coal, 0.8 percent sulfur
FGD vendor	ADL/LEA
Process	Lime/alkaline fly ash scrubbing
New or retrofit	New
Start-up date	July 1976
Efficiency, particulates	99 percent
Efficiency, SO_2	60 percent
Sludge disposal	Unstabilized sludge is disposed of in unlined pond

Each boiler is equipped with three scrubber modules, each module capable of handling 40 percent of the boiler flue gas capacity (1,430,000 acfm at 291°F). Each scrubber module incorporates a variable-throat venturi (design superficial gas velocity - 200 ft/sec), a spray absorption section (design superficial gas velocity - 11.5 ft/sec), an internal recycle tank, a wash tray, and a chevron-type mist eliminator. The venturi and spray sections operate on a recirculating slurry loop containing 12 percent solids (consisting primarily of calcium salts and fly ash). Total module pressure drop is approximately 23.5 inches H_2O. Total module liquid-to-gas ratio is approximately 33 gal/1000 acf. A wash tray is situated beneath the mist eliminator in order to dilute any entrainment reaching the component, thus reducing the potential for plugging and scaling. Slurry for this tray comes from a separate recycle tank and recirculation loop. Reheat is supplied by steam in heat exchanger tube bundles. The reheaters are also equipped with soot blowers for cleaning.

The scrubber booster fans (induced draft, 43.5 inches H_2O) are located downstream of the scrubbers and operate on dry, reheated flue gas. The fans were located downstream in this scrubber design because of the high particulate loading of the inlet flue gas.

The scrubbing wastes (unreacted fly ash, 9 parts calcium sulfate/1 part calcium sulfite) are discharged to an initial holding basin (520 acre-feet) adjacent to the scrubber plant and then pumped (50.8 dry tons per hour/2 units) 3 miles to a permanent sludge disposal pond (2218 acre-feet, clay-lined where necessary). Clear liquor is returned from ponds for additional scrubbing service. Fresh makeup water to the system is limited to approximately the amount of water lost in the scrubber plant via evaporation and sludge.

PERFORMANCE HISTORY

The Unit No. 1 FGD system was started up in late 1975. Following a shakedown and debugging period, the unit has been operating at full commercial capacity, and, according to the designers, approaching 100 percent availability. The Unit No. 2 FGD system started up in July 1976. The utility tested the units in 1976 and found both of them to be in compliance with the regulation calling for 1.2 lb $SO_2/10^6$ Btu. Some test runs indicated SO_2 removal efficiencies in excess of 90 percent.[6]

COLUMBUS AND SOUTHERN OHIO ELECTRIC CO.
CONESVILLE NO. 5

BACKGROUND

The Conesville Power Station is located on the Muskingum River, near Conshocton, in northeast Ohio. The plant has a current capacity of 1644 MW (design) with an additional 411 MW under construction. Units 1, 2, and 3 have a combined capacity of 433 MW and share a common stack. Unit 4 is rated at 800 MW, and Unit 5 is rated at 411 MW. Unit 6, currently under construction, will also be rated at 411 MW. Units 4, 5, and 6 each have a separate stack.

Boiler 5 is a dry-bottom, pulverized-coal-fired Combustion Engineering unit, installed in 1976. The plant burns a mixture of high-sulfur Ohio coals, with sulfur contents of from 4.2 to 5.1 percent, ash contents of from 12 to 19 percent, and heating values of from 10,300 to 11,220 Btu/lb. Forty percent of the coal is delivered by conveyor from a nearby (7 miles distant) coal mine complex. The remainder is trucked in from southeast Ohio.

POLLUTION CONTROL

The air pollution control system at Conesville No. 5 consists of a Research-Cottrell cold-side ESP, followed by two Universal Oil Products (UOP) SO_2 absorber modules in parallel. The ESP is designed for 99.65 percent removal efficiency and the Turbulent Contact Absorbers (TCA) are designed for 89.6 percent SO_2 removal efficiency. The system is designed for an outlet SO_2 loading of 1.0 lb/ million Btu of heat input. Boiler I.D. fans are located immediately downstream of the ESP.

Following the I.D. fans, the flue gas enters the two parallel TCA scrubbing trains. Each absorber is capable of handling 60 percent of the flue gas flow. A presaturator section lowers the flue gas temperature from 286° to 125°F, and provides some initial SO_2 removal. The gas then enters the neoprene-lined, carbon steel absorber modules, where two stages of 1.5-inch plastic balls provide a contacting surface between the lime slurry and the flue gas. Following each absorber module, the flue gas passes through a fiberglass entrainment separator and two horizontal banks of chevron-type mist eliminators. The bottom of the trap-out tray is washed intermittently and the lower mist eliminator is washed continually with recycled pond water. The flue gas from the parallel absorber trains then enters the 800-ft, Ceilcote-lined stack. Following the boiler I.D. fan is a bypass breeching around the entire scrubber loop. Each module can be bypassed independently. No stack gas reheat is currently being employed, although it is possible that reheat will be added on at some future date.

Dravo thiosorbic lime from Maysville, Kentucky, is utilized in the UOP scrubber modules at a stoichiometric ratio of 1.1. The calcined, pelletized lime has a nominal particle diameter of 1.75 inch, an MgO content of 3 to 8 percent and a CaO content of 90 to 95 percent. The lime slaker discharges the 20 percent solids slurry into an agitated lime slurry sump, where it is retained for a 5 minutes before being transferred to the lime slurry storage tank, which handles the surge requirements of the absorption system. The transfer of slurry from the storage tank to the TCA recycle tanks is accomplished by variable speed pumps, which respond to changing SO_2 concentrations and boiler load conditions via a pH monitor. The scrubbing liquor contains about 7 to 12 percent solids and is recirculated by four pumps (one standby), each rated at 12,000 gpm. Scrubber outlet pH is 5.8, and pH in the recycle tank is approximately 6.8.

A bleed stream of spent reaction products is continuously withdrawn from the recycle tank and pumped to the thickener system. The thickener is 100 ft in diameter and 14 ft deep in the center. Here the reaction product slurry is concentrated to an underflow composition of approximately 40 percent solids. This underflow is cycled to the IU Conversion Systems, Inc. (IUCS) fixation facilities where it is further thickened, vacuum-filtered, and mixed with a blend of dry fly ash and lime to form a 73 percent solid substance (IUCS Poz-o-tec). The product is currently being discharged to a 3500-acre-foot diked pond.

The wastewater pond receives ash sluice water, cooling water blowdown, and water from the sludge treatment plant. This system is not operating closed loop at the present time. Design information is presented in Table A-4.

PERFORMANCE HISTORY

The unit was originally scheduled to startup in December 1976. Prior to start-up, module 5A was destroyed by a fire in the neoprene lining. This module is being replaced and start-up is now scheduled for November 1977. The other module, No. 5B, began operation in January 1977. This was a very cold month and initial start-up problems were compounded by the severe weather. Frozen equipment and pipe-

Table A-4. FGD SYSTEM DATA, CONESVILLE NO. 5

Unit location	Conesville, Ohio
Unit rating	400 MW
Fuel characteristics	Coal, 4.5 to 4.9 percent sulfur
FGD vendor	Universal Oil Products
Process	Lime scrubbing
New or retrofit	New
Start-up date	February 1977
Efficiency, particulates	99.6 percent
Efficiency, SO_2	89.7 percent
Sludge disposal	Stabilized sludge is disposed of in unlined pond

lines were common. In the following months the facility continued to experience difficulties, including ruptured plastic piping, SO_2 analyzer failure, poor velocity distribution through the absorber, and carry-over of scrubber liquor into the mist eliminator. During shutdown in August, UOP replaced the pressurized plastic piping with stainless steel. Problems have included corrosion in the presaturator, failure of the rubber liner, and poor quality of the lime supply. The lime supplier is installing mechanical separators and metal detectors to improve the lime quality. The utility is considering retrofitting the unit with a reheat system.[7]

No SO_2 removal efficiency data are available at this time.

DUQUESNE LIGHT CO.
ELRAMA POWER STATION

BACKGROUND INFORMATION

The Elrama Power Station is owned and operated by the Duquesne Light Company. The station's four pulverized-coal-fired boilers have a generating capacity of 510 MW. Designed by Babcock and Wilcox, the boilers were placed into service in 1952, 1953, 1954, and 1960. Additional plant details can be found in Table A-5.[8]

POLLUTION CONTROL

The scrubber facility at the 510-MW Elrama Station is almost identical to the system at the Phillips Station,[+] also owned by Duquesne Light. Five Chemico single-stage venturi scrubbers have been installed after the mechanical collectors and electrostatic precipitators.

It was planned that the knowledge gained from the test program at the Phillips Station would be applied to Elrama to enable compliance with emission regulations.[9] The first Elrama scrubber was placed in service on October 26, 1975. It had been scheduled for an earlier start-up date, but because of the severity and number of problems encountered at Phillips, start-up was delayed until many of the problems at Phillips were resolved and the modifications could be incorporated at both stations.

Boiler No. 2 was initially connected to the FGD system on October 26, 1975. This boiler has an equivalent capacity

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[+] See description later in this Appendix.

Table A-5. FGD SYSTEM DATA, ELRAMA POWER STATION

Unit location	Elrama, Pennsylvania
Unit rating	510 MW
Fuel characteristics	Coal, 1.0 to 2.8 percent sulfur
FGD vendor	Chemico
Process	Lime scrubbing
New or retrofit	Retrofit
Start-up date	October 1975
FGD status	Operational
Efficiency, particulates	99 percent, design
Efficiency, SO_2	83 percent design - 2 percent sulfur coal
Sludge disposal	Stabilized sludge is disposed of in unlined pond

of approximately 100 MW, and the emissions are handled by one scrubber. To ensure reliability in the case of a scrubber malfunction, two scrubbers are operated, each at partial load.

The boiler operated continuously on the scrubber system through January 1976.* Two minor outages occurred during this period, one caused by inoperative throat dampers and the other by failure of a lime feeder belt.

Until additional construction is completed, two boilers are the maximum that can be served by the scrubber system.

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* Further operability data have not been reported due to recent litigation.

To date the utility has encountered the usual minor start-up problems in addition to some major problems. The major problems entailed frozen pipes and hardware and design problems associated with recirculation of the sludge within the thickeners to attain 30 to 40 percent solids concentration.

The single-stage scrubber trains were not designed for SO_2 removal, but through the use of hydrated calcium lime a 50 percent removal rate is being obtained in the interim period. Tests have shown that single-stage scrubbers can obtain at least 80 percent SO_2 removal by using a modified lime containing 6 to 10 percent MgO, which will eventually be used when all scrubber lime and sludge handling equipment is completed in early 1978.

ARIZONA PUBLIC SERVICE
FOUR CORNERS NO. 5

BACKGROUND INFORMATION

The Four Corners Power Station, located near Farmington, New Mexico, is jointly owned by the Arizona Public Service Company and the Southern California Edison Company. This station contains five coal-fired boilers with an equivalent electric generating capacity of 2233 MW.

Unit 5 is a coal-fired boiler with a net generating capacity of 800 MW. This unit was placed in commercial operation in July 1970. Approximately 20 percent of the flue gas generated by this boiler was scrubbed during a test program, to establish what sulfur dioxide and particulate removal efficiencies could be obtained.

POLLUTION CONTROL

The FGD system, which was retrofitted on Unit 5 in February 1976, is a Weir horizontal crossflow spray scrubber that was formerly installed and operated in a test program at the Mohave Generating Station of the Southern California Edison Company. Following completion of tests at Mohave, the module was dismantled, transported, and reassembled at the Four Corners Generating Station for further testing.

The horizontal module is a four-stage scrubbing chamber 48 feet long, 28 feet wide, and 15 feet high. The module itself is a hollow chamber consisting of four stages of 36 spray nozzles each. The slurry is cycled through the scrubber in a countercurrent manner at a rate of 36,000 gpm/stage. Fresh lime slurry is sprayed across the flue gas at the fourth stage, which is the discharge end of the scrubbing chamber. The scrubbing solution is then collected and recycled successively to the third, second, and first stages of the scrubbing module. This allows for more complete lime utilization and insures that fresh lime slurry contacts gas with the lowest sulfur dioxide concentration. The liquid recirculation rate can be adjusted over a wide range. The module is designed to treat a maximum flue gas capacity of 450,000 scfm (170 MW). The design gas velocity through the module is 21.6 ft/sec. The cleaned gases then pass through a demister and are reheated before being discharged to the stack. The spent slurry is discharged from the scrubbing system to a thickener and the underflow is pumped to an on-site settling pond. Water is returned to the process for further use.

The results of the test program, which was completed in December 1976, have not yet been released.[10]

KENTUCKY UTILITIES
GREEN RIVER BOILERS 1, 2 AND 3

BACKGROUND INFORMATION

The Green River Station of Kentucky Utilities is located on the Green River in central Kentucky, approximately 5 miles north of Central City. The local terrain surrounding the power plant is sparsely populated and heavily wooded.

The plant itself contains four steam turbine generating units with a total gross generating capacity of 242 MW. Boilers 1, 2, and 3 supply steam to generate 64 MW. These boilers are used for peak loads and normally operate on a 5-day week, with one or more of the boilers often at reduced load. All three boilers are dry-bottom, pulverized-coal-fired units manufactured by Babcock and Wilcox and installed in 1949 and 1950. There are presently no plans to retire these units. A coal with a 3.8 percent sulfur content is burned.

POLLUTION CONTROL

Boilers 1, 2, and 3 are fitted with mechanical collectors upstream of the FGD system. Design particulate removal efficiency is 85 percent. The FGD system was designed and installed by AAF and consists of a venturi scrubber and a single mobile-bed contactor module to handle a maximum flue gas capacity of 360,000 acfm at 300°F. Table A-6 summarizes the FGD system data.

Table A-6. FGD SYSTEM DATA,

GREEN RIVER NOS. 1, 2 AND 3

Unit location	Central City, Kentucky
Unit rating	64 MW
Fuel characteristics	Coal, 3.8 percent sulfur
FGD vendor	American Air Filter
Process	Lime scrubbing
New or retrofit	Retrofit
Start-up date	9/75
Efficiency, particulates	99.7 percent (design) (overall)
Efficiency, SO_2	80 percent (guaranteed)
Water makeup	1.20 gpm/MW
Sludge disposal	Sludge is disposed of in unlined pond

The flue gases from each boiler pass through a series of mechanical collectors where primary particulate removal takes place. The flue gas is then drawn from the existing breeching through a guillotine-type isolation damper and associated ductwork to the scrubber fan. The guillotine-type isolation dampers allow selective bypassing of the flue gas around the scrubbing system to the existing stack. The single-stage mobile-bed contactor consists of plastic balls through which the flue gas passes countercurrent to the lime slurry flow.

The scrubbing slurry feed and recycle system consists of a partitioned concrete reactant tank that includes recycle pumps to supply the scrubber and absorber module, a lime slurry slaking and feed system, and a system for discharging scrubbing waste to a settling pond. Water from the settling pond is recycled back to the scrubber.

PERFORMANCE HISTORY

This system has demonstrated good operability, as shown in Figure 3-23.[11] The problems encountered have been primarily of a general mechanical nature such as rubber peeling of the impellers. These problems were solved without significant scrubber down time.

Some chemistry-related problems also occurred. Specifically, a hard coat of scale developed in the lower sections of the absorber internals. This was believed to have occurred because of a high scrubbing solution pH combined with high oxygen concentrations in the flue gas. This problem was corrected by manually removing the scale, lowering the oxygen content of the flue gas by minimizing air leakage, and maintaining better pH control by modifying and relocating the pH sensors in the absorber. Following the completion of these modifications, scale formation has not been a problem although some scale film has been noted periodically, and has tended to disappear with continued operation.

In February 1977, the Carboline-lined stack was discovered to be badly deteriorated. The lining had failed in many areas, and the steel stack had corroded completely through in some areas.

The stack was repaired first by welding a backup metal plate to the portions of the stack where pitting occurred.

Then the entire stack was lined with refractory coating called Precrete G-8 by American Air Filter (AAF). The utility reports that the lining failure was due to poor application. Stack repairs were completed on March 7, 1977. Kentucky Utilities is now planning to install reheat on the system to prevent this problem from recurring. This stack liner problem resulted in the first delayed outage caused by the scrubber system.

SOUTHERN CALIFORNIA EDISON
MOHAVE

BACKGROUND INFORMATION

Two prototype sulfur dioxide absorber modules were installed in 1973 at the Mohave Generating Station of the Southern California Edison Company. A vertical module, rated at 170 MW, was installed to treat a 450,000-scfm portion of the flue gas from Unit 1. A horizontal module, also rated at 170 MW, was installed to treat a similar flue gas portion from Unit 2. Units 1 and 2 are identical boilers. Each has a maximum net continuous generating capacity of 790 MW, and each burns a western, low-sulfur coal with a heating value of 11,500 Btu/lb and ash and sulfur contents of 10 and 0.5 to 0.8 percent, respectively.

POLLUTION CONTROL

The vertical module, a Universal Oil Products turbulent contact absorber (TCA), was tested in two modes: first, as a TCA unit, from November 2, 1974, to April 30, 1975, and second, as a polygrid-packed absorber (PPA), from April 30, 1975 to July 2, 1975.

Start-up of the vertical module was initiated on schedule, January 1, 1974. On January 24, 1974, a fire burned most of the chlorobutyl lining, and the system was shut down until October 1, 1974. Testing of this module started a month later and was completed after 3131 hours of operation. System design data are presented in Table A-7. The module is presently shut down in a cold stand-by condition at the Mohave Generating Station. Refer to Section 3.1 of this report for detailed test data.

Table A-7. FGD SYSTEM DATA, MOHAVE NO. 1A

Unit location	Laughlin, Nevada
Unit rating	170 MW
Fuel characteristics	Coal, 0.5 to 0.8 percent sulfur
FGD vendor	Universal Oil Products
Process	Lime or limestone scrubbing
New or retrofit	Retrofit
Start-up date	November 1974
Efficiency, particulates	93 percent
SO_2 removal measured	95 percent
Water makeup	0.92 gpm/MW

Horizontal Module

The horizontal module, a four-stage horizontal cross-flow spray scrubber, was operated by the utility in a short series of start-up tests that ended on January 16, 1974, when a formal test program was initiated to assess the performance and reliability characteristics of the scrubbing system. The test program was terminated on February 9, 1975, following 5927 hours of operation. System design data are shown in Table A-8. The module was subsequently dismantled, transported, and reassembled at the Four Corners Plant of Arizona Public Service, where additional testing has been completed, but was not made available.

Test data are presented in Section 3.1 of this report.

Table A-8. FGD SYSTEM DATA, MOHAVE NO. 2A

Unit location	Laughlin, Nevada
Unit rating	170 MW
Fuel characteristics	Coal, 0.5 to 0.8 percent sulfur
FGD vendor	Southern California Edison/Stearns Roger
Process	Lime or limestone scrubbing
New or retrofit	Retrofit
Start-up date	November 1973
Efficiency, particulates	93 percent
SO_2 removal measured	95 percent
Water makeup	1.0 gpm/MW
Sludge disposal	Stabilized sludge is disposed of in unlined pond

LOUISVILLE GAS AND ELECTRIC CO.

PADDY'S RUN NO. 6

BACKGROUND INFORMATION

The Paddy's Run Power Station of Louisville Gas and Electric Company is located on the Ohio River in Rubbertown, about 10 miles southwest of the center of Louisville, Kentucky. Of the six units at Paddy's Run, only the boiler on Unit 6 is retrofitted with an FGD system.

This boiler is a dry-bottom, pulverized-coal-fired unit designed and installed by Foster-Wheeler in 1951. It operates as a peaking unit and has a design rating equivalent to 65 MW. The station as a whole operated at only an approximate 5 percent load factor in 1974. Because of this low load factor, the other boilers are not being fitted with an FGD system. Data on plant operation and emissions appear in Table A-9.

The coal now being burned has an average heating value (as received) of 12,400 Btu/lb. Ash and sulfur contents are 14 and 3.5 to 4 percent, respectively.

POLLUTION CONTROL

A Research-Cottrell electrostatic precipitator (ESP) with an efficiency of 99 percent provides primary control of particulate emissions. Particulate loading at the outlet of the ESP unit is approximately 0.05 gr/scf.

Table A-9. FGD SYSTEM DATA, PADDY'S RUN NO. 6

Unit location	Louisville, Kentucky
Unit rating	65 MW
Fuel characteristics	Coal, 3.5 to 4.0 percent sulfur
FGD vendor	Combustion Engineering
Process	Lime scrubbing
New or retrofit	Retrofit
Start-up date	April 1973
Efficiency, particulates	99 percent (design)
Efficiency, SO_2	80 percent (design)
Water makeup	Open-loop, 0.7 gpm/MW
Sludge disposal	Unstabilized sludge is disposed of in unlined pond

Atmospheric emissions of sulfur dioxide are limited to 1.2 lb/MM Btu of heat input, and continuous monitoring equipment shows that SO_2 emissions are within this limit.

The FGD system was designed by Combustion Engineering, Inc. Start-up for the FGD plant occurred in April 1973. The lime scrubbing system utilizes calcium hydroxide sludge, which is generated as a waste by-product at a nearby acetylene manufacturing plant. In this FGD process a slurried mixture of calcium hydroxide and calcium sulfite in water constitutes the scrubbing liquor.

The FGD system consists of two identical modules, each sized to handle 175,000 acfm of flue gas at 350°F.

Table A-9. FGD SYSTEM DATA, PADDY'S RUN NO. 6

Unit location	Louisville, Kentucky
Unit rating	65 MW
Fuel characteristics	Coal, 3.5 to 4.0 percent sulfur
FGD vendor	Combustion Engineering
Process	Lime scrubbing
New or retrofit	Retrofit
Start-up date	April 1973
Efficiency, particulates	99 percent (design)
Efficiency, SO_2	80 percent (design)
Water makeup	Open-loop, 0.7 gpm/MW
Sludge disposal	Unstabilized sludge is disposed of in unlined pond

Atmospheric emissions of sulfur dioxide are limited to 1.2 lb/MM Btu of heat input, and continuous monitoring equipment shows that SO_2 emissions are within this limit.

The FGD system was designed by Combustion Engineering, Inc. Start-up for the FGD plant occurred in April 1973. The lime scrubbing system utilizes calcium hydroxide sludge, which is generated as a waste by-product at a nearby acetylene manufacturing plant. In this FGD process a slurried mixture of calcium hydroxide and calcium sulfite in water constitutes the scrubbing liquor.

The FGD system consists of two identical modules, each sized to handle 175,000 acfm of flue gas at 350°F.

Flue gas enters each scrubber module near the base and contacts nonatomizing sprays which provide a constant supply of slurry to the underside of the two stages of marble beds. After emerging from the second marble bed, the clean flue gas passes through a two-stage chevron mist eliminator, where entrained water droplets are agglomerated and removed. The flue gas then passes through a gas-fired reheater, a booster fan, and out the stack.

Scale formation does not occur at this plant as long as both of the following conditions are met: 1) fly ash must not constitute more than about 6 percent of the slurry solids, and 2) slurry pH in the reaction tank must be maintained between 8.0 and 9.5. Slurry pH is measured downstream from the reaction tank. Additive lime is pumped to the reaction tank from the additive slurry tank. Scaling potential does not seem to be influenced by the oxygen content of the exhaust gas, which typically ranges between 6 and 9 percent.

The solids content of the slurry leaving the absorber ranges between 9.5 and 10.5 percent. The ratio of sulfite to sulfate in the recirculating slurry ranges between 40 and 50:1 on a weight basis. Liquid-to-gas ratio (L/G) ranges between 15 and 18 gal/1000 ft^3 of gas at 125°F per stage. Fresh slurry additive contains 20 to 30 percent solids and has a pH of 12.6. Lime is added to the thickener tank to stabilize the sludge that is formed at a rate of about 100 pounds of lime per ton of dry sludge. The sludge is trucked to a 10-acre landfill that ranges from 20 to 30 feet in depth. Solids content of the effluent from the thickener averages about 25 percent. This material is then dewatered by vacuum filtration to form a stabilized sludge containing

45 percent solids. Alternately, the thickener effluent can be mixed with dry fly ash and lime to form stabilized sludge with a 45 percent solids content.

No major start-up problems were encountered that were associated with the scrubbers. Problems with other system components included the following:

Thickener Size - The original thickener was too small to handle the untreated slurry at full load. Preliminary data had indicated that the slurry material entering the thickener would contain more calcium sulfate than was actually encountered. The slurry actually contained a high ratio of calcium sulfite to calcium sulfate.

Carbide Lime Feed System - Some early problems were experienced with plugging of the mesh strainer on the additive tank. This problem was solved by installing a Rietz mechanical disintegrator in the carbide feed line to the additive tank.

Mist Eliminator Wash System - The original mist eliminator wash system consisted of rotating nonretractable lances with 3/16-inch diameter nozzles every 6 inches. The system would not efficiently clean the upper mist eliminator. The nonretractable lances were replaced with retractable half-track lances with oscillating 1/2-inch nozzles. The larger solid stream of water from these nozzles keeps the mist eliminators clean.

Damper Leakage - The FGD system can be bypassed by means of louvered dampers. Deposits on these dampers were found to prevent them from closing completely, so leakage through the dampers resulted in the formation of a visible plume. By cleaning the dampers periodically the leakage was minimized and plume formation effectively suppressed.[12]

PERFORMANCE

In July 1977, FGD efficiency tests were run with the addition of MgO. Results from those two 8-hour test runs are as follows:

Date	SO_2 concentration, ppm Inlet	 Outlet	Removal, %
July 8, 1977	2230	1 to 5	99.7-99.9
July 16, 1977	2150	1 to 5	99.7-99.9

Additional test data are presented in Section 3.1 of this report.

Availability figures approaching 100 percent have been achieved during recent operation, but the power plant, which is primarily a summer peaking unit, is operated infrequently. The capacity factor for Boiler 6 in 1976 was about 14.6 percent. The FGD unit was operated only 7 months in 1976.[13]

DUQUESNE LIGHT CO.
PHILLIPS POWER STATION NO. 6

BACKGROUND INFORMATION

The Phillips Power Station of the Duquesne Light Company is located on the Ohio River in Allegheny County, Pennsylvania, 20 miles northwest of Pittsburgh, Pennsylvania. The area is heavily industrialized, and the plant supplies electricity to residential, commercial, and industrial users in Allegheny and Beaver Counties.

Six generating units at this station constitute a gross continuous generating capacity of 408 MW. The net station capacity is 373 MW when all four particulate scrubbers and the single SO_2 absorber are operating. All the boilers are dry-bottom, pulverized-coal-fired units manufactured by Foster-Wheeler. The first unit was installed in 1942 and the sixth one in 1956. Coal at this station has an average as-received gross heating value of 11,350 Btu/lb. Ash content, on a dry basis, is 18.2 percent, and sulfur content is 2.15 percent.[14] All the generators are cycling base load units. Unit 6 is the largest boiler, with a net capacity of 143 MW.

POLLUTION CONTROL

Primary particulate emission control at this plant is provided by Research-Cottrell mechanical collectors followed by a Research-Cottrell electrostatic precipitator on each boiler. The collection efficiency for this system was originally estimated to be 90 percent; however, the addition of the FGD system caused disruption of the gas flow distribution through the ESP, and the combined efficiency through the mechanical and electrostatic system now appears to be only about 80 percent.

Final particulate control is achieved by four parallel Chemico venturi scrubbers with a design efficiency of 99 percent. These scrubbers also remove about 50 percent of the SO_2 when lime is added to the water. FGD data are presented in Table A-10.

Table A-10. FGD SYSTEM DATA, PHILLIPS POWER STATION NO. 6

Unit location	South Height, Pennsylvania
Unit rating	143 MW
Fuel characteristics	Coal, 1.0 to 2.8 percent sulfur
FGD vendor	Chemico
Process	Lime scrubbing
New or retrofit	Retrofit
Start-up date	July 1973
Efficiency, particulates	99 percent (design)
Efficiency, SO_2	83 percent (design, 2 percent sulfur coal)
Sludge disposal	Stabilized sludge is disposed of in unlined pond

Emissions from one of the particulate scrubbers are further controlled by a Chemico second-stage venturi absorber utilizing a lime solution. Sulfur dioxide removal efficiency for this two-stage train has averaged about 90 percent, based on performance guarantee tests. Sulfur dioxide removal efficiency of the single-stage scrubber is 50 to 60 percent when the pH is maintained between 6 and 7. By December 1977, through the use of a single-stage scrubber and the firing of coal with 2 percent sulfur content, the unit should be able to achieve 83 percent SO_2 removal when thiosorbic lime (6 to 12% Mg) is used as the absorbent.

PERFORMANCE

The Phillips Station was the first lime FGD system in the United States. The Phillips scrubber system was started up in July 1973. Several problems developed in the form of erosion of fans and the lime feed system, causing outages of the scrubber system. After an extended outage, the scrubber system was returned to service in March 1974. Scale deposits were removed, corroded parts were replaced, and "off-spec" components were replaced with the originally specified material. Not all of the operating problems have been solved, and system availability has been lower than that of many other FGD systems. Scale continues to be a problem, and the utility is planning to use lime with magnesium oxide added in an attempt to reduce the problem. Sometimes the FGD system has been bypassed so that outages would not reduce boiler output. Problem areas are discussed below:

Recycle Pumps - The original Carpenter 20 alloy impellers and casings used in the Ingersoll-Rand scrubber pumps have failed repeatedly.

Approximately 30 percent of the metal on both impellers and casings is being eroded/corroded, thus limiting pump service life to about 1000 to 3000 hours. An extensive pump reliability study is under way. Several materials of construction are being tried, including rubber and hard iron.

Throat Dampers - Throat dampers have been a major problem because of fly ash and hard scale deposits. The deposits caused an increase in pressure drop across the throat from 6 inches to 12 inches or more. This increase in pressure drop reduced the amount of gas that could be scrubbed. In an attempt to clear the deposits, Duquesne Light maintained a continuous flow of water through the 3/4-inch openings of the damper housing. When this effort failed, air was forced through an open vent on the sides of the damper housings. The problem remains unsolved. Significant accumulations are still being found in the throat damper. Manual cleaning requires removal of the train from service and takes about 288 man-hours per scrubber.

A total of 60 throat damper arms (12 per vessel) were specified as 316L SS, but were apparently fabricated from 304 SS. Thirty-two of these throat damper arms corroded badly. The top and bottom throat damper scraper blades were also found to be 304 SS; they were essentially destroyed by corrosion. All of the damaged throat damper arms and damper scraper blades were replaced with 316L SS components.

Wet Induced Draft Fans - Several problems have occurred in operation of the induced draft fans. The fan shaft shrouds, the 316 SS-clad fan dampers, and the stiffener bars have all corroded. The fan spray sump pumps have also been wearing out much faster than anticipated, partly because of a chloride concentration of about 3500 to 4500 ppm and a comparatively low pH in the scrubber liquor. Several steps were taken to alleviate the problems: the fan spray pumps were redesigned; the affected fan shaft shroud parts were coated with 1/4-inch neoprene rubber; addition of lime to the system was increased to achieve optimum pH; double (reinforcing) plates were installed on each of the blades to reduce the fan stresses causing metal deformation; and deposits were manually removed from the fans. Despite all of these steps, significant pitting is still taking place on the back of the fan blades, primarily because of fly ash/sludge deposits.

Lime Slurry Preparation - The lime slaker at the Phillips plant was not designed for use with all six boilers connected to the system. Discontinuous operation of the slaker system has resulted in deviations from proper pH and has caused deposition of solids.[15]

U.S. AIR FORCE
RICKENBACKER AIR FORCE BASE

BACKGROUND INFORMATION

The FGD system at the Rickenbacker Air Force Base located near Columbus, Ohio, is a Research-Cottrell/Bahco System which was started up in March 1976.

This first U.S. Bahco installation handles the flue gas from seven stoker-coal-fired boilers. Three boilers were first placed in service in 1952, and the remaining four in 1954. One is rated at 31 million Btu/hr and the other six at 60 million Btu/hr each. The units fire a high-sulfur (3.6 percent) coal at high combustion air rates (150 to 160 percent excess air). The plant burns approximately 40,000 tons of coal per year. Plant and FGD data are presented in Table A-11.[16] As shown in the table, the FGD system is designed to remove between 82 and 98 percent of the SO_2.

POLLUTION CONTROL

The Rickenbacker emission control system consists of a mechanical collector, Bahco scrubber tower, lime storage and handling system, clarifier, booster fan and associated ductwork, pumps, controls, and a sludge disposal pond.

Particulate and SO_2-laden gas taken from each of the existing seven stacks is fed through a common header and into a mechanical collector, where primary particulate removal takes place. The mechanical collector has a design removal efficiency of 70 percent and was installed primarily to reduce fan wear. The partly cleaned gas then flows to the fan and into the Bahco tower, where additional particulate and primary SO_2 removal takes place.

Table A-11. FGD SYSTEM DATA, RICKENBACKER AIR FORCE BASE

Unit location	Columbus, Ohio
Unit rating	20 MW
Fuel characteristics	Coal, 3.6 percent sulfur
FGD vendor	Research-Cottrell
Process	Research-Cottrell/Bahco Lime Scrubbing
New or retrofit	Retrofit
Efficiency, particulate	98 percent
Efficiency, SO_2	90 percent (82 to 98% depending on stoichiometric lime feed rate)
Sludge disposal	Unstabilized $CaSO_3/SO_4$ sludge to lined pond

The Bahco scrubbing vessel is a vertical tower consisting of two converted venturi stages. Gas flows through the tower, where it is contacted with a lime slurry solution. After passing through the first scrubbing stage, entrained moisture is removed by a cyclonic-type mist eliminator. The gas then passes through a second scrubbing stage and a cyclonic-type mist eliminator. The cleaned gas is discharged to the atmosphere through a stack mounted directly on the tower.

Pebble lime is used as the reagent in the scrubbing system and is prepared by slaking and dissolution in a mix/

hold tank. The scrubbing slurry is then pumped to the absorber, where it is contacted with the flue gas in a countercurrent fashion.

The spent scrubbing solution is discharged to the thickener, where the waste solids settle out. Thickener overflow is returned to the mix/hold tank. Thickener underflow is discharged to a 5-acre Hypalon-lined disposal pond. This pond is located approximately 400 feet from the scrubbing tower. Clear water from the pond is returned to process as slaking water.[17]

Research-Cottrell has recently released data that indicates the SO_2 removal efficiency can approach 100 percent if the stoichiometric ratio exceeds 1.0.[18]

TENNESSEE VALLEY AUTHORITY
SHAWNEE NO. 10A AND 10B

BACKGROUND INFORMATION

The Shawnee Steam Plant is located on the Ohio River about 10 miles northwest of Paducah, Kentucky. The plant has 10 separate pulverized-coal-fired boilers, each connected to its own stack and generating turbine. The boilers, which were manufactured by Babcock & Wilcox, are identical in design, each rated at 175 MW capacity. The first unit went into commercial operation in April 1953. By June 1957, all 10 units were generating power.

POLLUTION CONTROL

Particulate emissions from each boiler are controlled by a mechanical collector followed by an electrostatic precipitator. An experimental FGD prototype plant, jointly sponsored by EPA and TVA, is being operated on a portion of the flue gases from Unit 10. Data from this system are presented in Tables A-12 and A-13.

The two parallel, retrofitted FGD systems began operation in April 1972. Universal Oil Products manufactured one of the FGD systems, a turbulent contact absorber. Chemical Construction Company (Chemico) produced the other FGD system, which consists of a venturi scrubber followed by a spray tower. Each FGD system is capable of treating 30,000 acfm, the equivalent of approximately 10 MW of power plant generating capacity. A third (experimental) FGD system, manufactured by Combustion Engineering, Inc., was operated for a time at Shawnee, but testing of this system has been concluded.

In June 1974, the EPA began a 3-year advanced test program at the Shawnee facility. The major goals are as follows: (1) continuation of long-term testing, with emphasis placed upon the reliable operation of mist elimination systems; (2) investigation of advanced process and equipment design variations for improving system reliability and process economics; and (3) long-term reliability testing on promising process and equipment design variations. The testing program is scheduled to run through June 1978. Because of the experimental nature of the FGD systems, no system operability data are published.[19] The system has operated well, however, and yielded useful data. Operating data and test results are presented in Section 3.1.

Many of the operational problems of FGD systems have been solved at this facility. Some of the most important findings are discussed below. Mist eliminator scaling was controlled by using a top wash with fresh water every 4 to 8 hours when alkali utilization was in excess of 85 percent. The benefits and techniques of forced oxidation were also studied, because they are related to alkali utilization. The effects of magnesium liquid-to-gas ratio, chloride, pH, and stoichiometry on SO_2 removal were studies. Turndown capabilities were demonstrated for the venturi/spray system.[20]

Table A-12. FGD SYSTEM DATA SHAWNEE NO. 10A

Unit location	Paducah, Kentucky
Unit rating	10 MW
Fuel characteristics	Coal, 2.9 percent sulfur
FGD vendor	Universal oil products
Process	Lime/limestone scrubbing
New or retrofit	Retrofit
Start-up date	April 1972
Removal Efficiency, SO_2	60 to 98 percent

Table A-13. FGD SYSTEM DATA SHAWNEE NO. 10B

Unit location	Paducah, Kentucky
Unit rating	10 MW
Fuel characteristics	Coal, 2.9 percent sulfur
FGD vendor	Chemico
Process	Lime/limestone scrubbing
New or retrofit	Retrofit
Start-up date	April 1972
Removal Efficiency, SO_2	60 to 90 percent

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APPENDIX B
DOMESTIC LIMESTONE SLURRY FGD SCRUBBING SYSTEMS

This appendix summarizes information on limestone scrubbing FGD systems presented in Section 3.2 and also describes other operating limestone scrubbing systems in this country.

The following FGD systems are described:

- Arizona Public Service - Cholla No. 1
- Kansas City Power & Light - La Cygne No. 1
- Kansas City Power & Light - Lawrence Nos. 4 and 5
- Northern States Power Co.
- Springfield Utilities
- Widows Creek No. 8
- South Carolina Public Service Co.
- Will County No. 1

The EPA test facility at Shwnee uses both lime and limestone. It is described in Sections 3.1 and 3.2 and in Appendix A.

ARIZONA PUBLIC SERVICE
CHOLLA NO. 1

BACKGROUND INFORMATION

The Cholla Power Generating Station is located near Joseph City, in Navajo County, Arizona. At present Cholla operates only Unit 1, a dry-bottom, pulverized-coal-fired boiler with a net generating capacity of 115 MW. The boiler was designed by Combustion Engineering, Inc. Two 250-MW boilers are under construction and two 350-MW boilers are also being planned at the Cholla Power Station.

The coal now being burned has typical fuel values, as received, of 10,290 Btu/lb, 10.4 percent ash and 0.4 to 1.0 percent sulfur. It is fired at a rate of 54 tons/hr at capacity.

POLLUTION CONTROL

A Research-Cottrell multicyclone-type collector, operating with an efficiency of about 75 percent, provides primary control of particulate emissions. Design particulate loading at the outlet of the multicyclones is approximately 2 gr/scf.

In 1973, the boiler was retrofitted with a flooded-disc scrubber system for particulate and SO_2 control. Designed by Research-Cottrell, Inc. (R-C), the wet limestone system removes 99.7 percent of the particulate matter and 92 percent of the SO_2.* The flue gas desulfurization system consists of two parallel scrubbing modules, each designed to accommodate 50 percent of the flue gas. Module A of the system includes an adjustable flooded-disc scrubber for particulate control, followed by a packed tower, which utilizes a limestone slurry, for SO_2 removal. Module B also incorporates a flooded-disc scrubber for particulate control but does not currently utilize a second stage. Sulfur dioxide removal efficiency for Module B is 25 percent, whereas that for Module A is 92 percent. FGD data are summarized in Table B-1.

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* Average SO_2 removal efficiency for the SO_2 and particulate scrubber is 58.5 percent (92 percent for A-side and 25 percent for B-side, 92+25/2 = 58.5).

Table B-1. FGD SYSTEM DATA, CHOLLA NO. 1

Unit location	Joseph City, Arizona
Unit rating	115 MW
Fuel characteristics	Coal, 0.44 to 1 percent sulfur
FGD vendor	Research-Cottrell
Process	Limestone scrubbing
New or retrofit	Retrofit
Start-up date	October 1973
Efficiency, particulates	99.7 percent
Efficiency, SO_2 (actual)	58.5 percent overall 92 percent in absorber
Water makeup	Open-loop, 1.04 gpm/MW
Sludge disposal	Unstabilized sludge is disposed of in unlined pond

PERFORMANCE HISTORY

Although the FGD system at Cholla has a good operating history, operation has not been trouble-free. Reheater vibration was an early problem, caused by a difference in the size of the main duct and the reheater transition duct. Baffles were installed to eliminate this problem and to prevent acid condensation on the reheater tubes, where corrosion was occurring.

Solids buildup and line plugging occurred, particularly when the system operated at low flow rates. To solve these problems, the piping was modified to eliminate stagnant pockets where solids could accumulate, and pumps were flushed immediately after removal from service.

Overall operability has been inconsistent, with many periods of 100 percent and a low period of 46 percent.

The overall SO_2 removal efficiency is 58.5 percent, which is sufficient to meet the local regulations.

FUTURE PLANS

Arizona Public Service (APS) plans to install an FGD system on Cholla Unit 2, which is presently under construction. APS signed a contract with R-C to purchase a particulate scrubber and an FGD system for that boiler similar to the system on Unit 1. R-C is preparing preliminary engineering designs and plans to use two modules to treat the flue gases from that boiler.*[1,2,3]

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* Personal communication from Mr. Lyman K. Mundth, Arizona Public Service Co. to Mr. Don R. Goodwin, U.S. Environmental Protection Agency. July 20, 1977.

KANSAS CITY POWER & LIGHT
LA CYGNE NO. 1

BACKGROUND INFORMATION

The La Cygne Power Station of Kansas City Power and Light (KCP&L) is a new station located about 55 miles south of Kansas City, in Linn County, Kansas.

The electric power generating facilities consist of one Babcock & Wilcox 6,200,000 lb steam/hr, coal-fired, base-load boiler with associated 820-MW (net) steam turbine and electric generator. The plant also has three oil-fired boilers, used primarily for start-up of the large unit and to supply steam to a separate 22-MW turbine generator.

The coal now being burned ranges in gross heating value (as received) from 8200 to 10,200 Btu per pound. Ash and sulfur range from 20 to 30 percent and from 5 to 6 percent, respectively.

POLLUTION CONTROL

The FGD system consists of eight particulate and SO_2 scrubbing modules, with on-site limestone grinding and storage facilities. All flue gases are treated, and the ductwork does not provide for the bypassing of flue gas around the scrubbers. As the hot flue gas enters the venturi, it is subjected to jets of limestone slurry injected through nozzles on the walls of the vessel. The liquid-gas stream flows downward through the venturi throat, where the gas contacts the atomized liquid droplets. The scrubbing efficiency is regulated by adjusting the venturi throat gap. As the gas exits from the venturi and enters the disengagement chamber, its velocity decreases from about 130 ft/sec (at the throat) to about 15 ft/sec. This reduction in velocity separates the limestone slurry droplets from the quenched gas. The slurry drains into the recirculation tank. The gas then enters the SO_2 absorber tower base and moves upward through two sieve trays in series. As the gas passes through the 1.375-inch-diameter holes in the sieve trays, it contacts limestone slurry that is sprayed into the path of the rising gas. The scrubbed gas then passes through a third sieve tray, which collects slurry carryover and reduces the mist load on the demister. The gas then passes through a 10-inch-high chevron demister, where the remaining fine droplets coalesce and drip back down through the gas stream into the recirculation tank. The flue gas is then reheated from about 121° to 175°F. Reheating is accomplished primarily by means of steam coils, with additional heat provided by injecting hot air from the boiler combustion air heater. Additional steam reheat capacity is currently being installed to eliminate the need for hot air bypass.

Additional FGD system data are presented in Table B-2.

PERFORMANCE HISTORY

The FGD system has encountered typical problems. Some start-up problems, such as vibrations of the I.D. fans and the sensitivity of these fans to imbalance, occurred even before the boiler was fired and are not related to the FGD system. As these fabrication problems were corrected, other types of problems appeared. These included plugging of the

Table B-2. FGD SYSTEM DATA, LA CYGNE NO. 1[4]

Unit location	La Cygne, Kansas
Unit rating	820 MW (net)
Fuel characteristics	Coal, 5.0 percent sulfur
FGD vendor	Babcock & Wilcox
Process	Limestone scrubbing
New or retrofit	New
Start-up date	February 1973
Efficiency, particulates (actual)	98 percent
Efficiency, SO_2 (actual)	80.14 percent
Water makeup	Open-loop, 1.4 gpm/MW
Sludge disposal	Unstabilized sludge is disposed of in unlined pond

demister and strainers, wearing of spray nozzles, and corrosion of reheater tubes. Problems of this type are the result of collected fly ash and flow characteristics of the recirculated slurry. Several modifications have been made to the original plant design in an effort to solve these problems. These include hydroclones to separate fly ash, increased reheat capacity, and installation of a spare scrubber module.

Since there is no bypass, a considerable amount of effort is expended toward maintaining the modules by nightly cleaning of one module per night.

This system has shown continued performance improvement as shown in Figure 3-28. Availability for the last 18 months has been in excess of 90 percent.

The results of a stack sampling test conducted on May 5, 1975, with the unit continuously operating from 700 to 720 MW, indicated an average SO_2 removal efficiency of 80.14 percent, and removal of 98.2 percent of the particulates.

SO_2 removal efficiency is 80.14 percent, which is sufficient to meet local regulations.[4,5,6,7]

KANSAS POWER & LIGHT COMPANY
LAWRENCE NO. 4 AND NO. 5

BACKGROUND INFORMATION

The Lawrence Power Station of Kansas Power and Light Company is located in a lightly industrialized area on the outskirts of Lawrence, Kansas. The plant operates two Combustion Engineering steam boilers, which are equipped to burn coal or natural gas supplemented with oil, or a combination of these three fuels. Boiler 4, the older of the two units, was first placed in service in 1959. It is operated as a cyclic load boiler. The maximum electric generating capacity of this unit varies with the type of fuel being burned; output can be as high as 143 MW when burning natural gas, whereas it decreases to 125 MW when burning coal plus natural gas. The retrofitting of this boiler with an FGD system in 1968 reduced the boiler capacity to 115 MW.

The second unit at the plant is Boiler 5. Its rated capacity, when burning coal plus natural gas, is 400 MW. The unit and the FGD system were placed in service in November 1971. Boiler 5 is similar to Boiler 4, and it also is classified as a cyclic load unit.

The company has now switched from high-sulfur Kansas coal to Wyoming coal, which contains from 0.4 to 0.8 percent sulfur and 10 percent ash. The coal has a gross heating value of 10,000 Btu/lb. Tables B-3 and B-4 present a summary of the FGD systems.

Table B-3. FGD SYSTEM DATA, LAWRENCE NO. 4

Unit location	Lawrence, Kansas
Unit rating	125 MW
Fuel characteristics	Coal, 0.4-0.8 percent sulfur
FGD vendor	Combustion Engineering
Process	Limestone scrubbing
New or retrofit	Retrofit
Start-up date	December 1968
Efficiency, particulates	99 percent
Efficiency, SO_2	75 percent
Sludge disposal	Unstabilized sludge is disposed of in unlined pond

Table B-4. FGD SYSTEM DATA, LAWRENCE NO. 5

Unit location	Lawrence, Kansas
Unit rating	400 MW
Fuel characteristics	Coal, 0.4-0.8 percent sulfur
FGD vendor	Combustion Engineering
Process	Limestone scrubbing
New or retrofit	New
Start-up date	November 1971
Efficiency, particulates	99 percent
Efficiency, SO_2	65 percent
Sludge disposal	Unstabilized sludge is disposed of in unlined pond

POLLUTION CONTROL

Unit 4 was originally a dry limestone injection system, but it has been converted to a limestone flue gas scrubbing system. Boiler 4 has two FGD modules, and Boiler 5 has eight. All are identical in size, and each is designed to handle approximately 150,000 scfm of flue gas. Originally, each module consisted of a single stage of 3/4-inch glass marbles. The bed was about 3 to 4 inches thick and was fitted with overflow pots to collect the liquor drained from the top of the bed. The scrubbing liquor was sprayed through nozzles located below the bed. The utility is in the process of replacing the marble bed absorbers with spray towers. Boiler 4 was completed in 1977; Boiler 5 is scheduled for completion in 1978.

Each of the two modules on Boiler 4 is connected (through an I.D. fan) to a separate 120-ft stack, whereas the gases from all eight modules on Boiler 5 are discharged through a common 375-ft stack.

PERFORMANCE HISTORY

Availability figures have not been reported; however, the availability is apparently adequate for this station, since it has a low load factor and up to 50 percent of the modules can be shut down for maintenance. No lost boiler capacity has been reported on Unit 4 since January 1977, when the limestone scrubbing spray towers were started.[8,9,10]

Measured SO_2 efficiency has not been reported.

NORTHERN STATES POWER CO.
SHERBURNE COUNTY STATION NO. 1 AND NO. 2

BACKGROUND INFORMATION

The Sherburne County generating plant of the Northern States Power Company is located adjacent to the Mississippi River in Sherburne County, near Becker, Minnesota. Each of the two units has a net capability of 700 MW. The boilers are controlled-circulation, single-reheat, balanced-draft units manufactured by Combustion Engineering. Two additional units are being planned to boost the total plant output to 3000 MW by 1980.

The coal presently in use is subbituminous western coal. It contains approximately 28 percent moisture, 9 percent ash, 0.8 percent sulfur and 8300 Btu/lb.

POLLUTION CONTROL

The scrubber system utilizes wet limestone scrubbing for both particulate and SO_2 removal. Twelve scrubber modules are provided for each unit with only 11 required for full load operation. Design particulate and SO_2 removal efficiencies of these systems are 99 percent and 50 percent, respectively, as shown in Tables B-5 and B-6.

The Combustion Engineering scrubber systems consist of a rod-type venturi throat, a marble bed absorber, a two-stage fiberglass Chevron demister, and a fine tube reheater.

PERFORMANCE HISTORY

Problems encountered by Sherburne include some mechanical problems, partly due to a last-minute decision to include a particulate scrubber. Mud deposits in and around the venturi throat and erosion of the walls and rods in the throat were a problem. These areas have been relined with 316L stainless steel. The components are still wearing, but at a slower rate.

Other problems are loss of marbles with the sludge, reheater leaks, failure of rubber lining in pipes, erosion of pump internals, ineffectiveness of Zurn duplex strainer, and limited capacity of limestone grinding mills. All of these problems are being dealt with either by design modification such as 316 stainless wear plates or by replacement parts.

Availability for Unit 1 averaged 85 percent for the first 4 months of operation after start-up. For the past 12 months, availability has been in excess of 90 percent. Unit 2 has shown even better start-up performance, with operability averaging about 95 percent for the first 4 months. These data are shown in Figure 3-29.

Unit 1

1977 - The unit was down until June 6 for completion of the first year's inspection. It came back on line June 18, 1977.

FUTURE PLANS

Northern States Power and C-E are sponsoring a long-term optimization program for the units.[11,12,13]

Table B-5. FGD SYSTEM DATA, SHERBURNE COUNTY STATION NO. 1

Unit location	Becker, Minnesota
Unit rating	710 MW
Fuel characteristics	Coal, 0.8 percent sulfur
FGD vendor	Combustion Engineering
Process	Limestone scrubbing
New or retrofit	New
Start-up date	March 1976
Efficiency, particulates	99 percent (guaranteed)
Efficiency, SO_2	50 percent (guaranteed)
Water makeup	Open-loop, 1.13 gpm/MW
Sludge disposal	Unstabilized sludge is disposed of in lined pond

Table B-6. FGD SYSTEM DATA, SHERBURNE COUNTY STATION NO. 5

Unit location	Becker, Minnesota
Unit rating	680 MW
Fuel characteristics	Coal, 0.8 percent sulfur
FGD vendor	Combustion Engineering
Process	Limestone scrubbing
New or retrofit	New
Start-up date	April 1977
Efficiency, particulates	99 percent (guaranteed)
Efficiency, SO_2	50 percent (guaranteed)
Water makeup	Open-loop, 1.13 gpm/MW
Sludge disposal	Unstabilized sludge is disposed of in lined pond

SPRINGFIELD CITY UTILITIES
SOUTHWEST NO. 1

BACKGROUND INFORMATION

The Southwest Power Station of Springfield City Utilities is located in Springfield Missouri. The plant operates one coal-fired boiler (Unit 1) with a rating of 200 MW. The boiler uses coal with a 3.5 percent sulfur content.

POLLUTION CONTROL

The two turbulent contact absorber modules at the Southwest Power Station are supplied by Universal Oil Products. These limestone scrubbers are downstream of an ESP and are designed for overall removal of 99.7 percent of the particulate emissions and 80 percent of the SO_2 emissions. This unit, which was put on-line in April 1977, has experienced some start-up problems. Specifically, these have included mist eliminator plugging, corrosion in the presaturator chamber, corrosion of the inlet/outlet gas dampers, and failure of the lining materials in both the scrubber and stack. These problems are being solved by a number of modifications:

- Modification of the scrubber outlet damper
- Replacement of scrubber packing
- Relining of the scrubber outlet ductwork
- Insulating of the inlet dampers

Sludge produced by the FGD system is deposited in an unlined pond after stabilization. FGD system data are summarized in Table B-7.

FUTURE PLANS

One module is being taken off line to implement modifications while the other remains on-line. If the modifications are successful, then the other module will also be modified.

PERFORMANCE HISTORY

Figures on operability and actual SO_2 removal efficiency are not yet available on this FGD system.[14,15,16]

Table B-7. FGD SYSTEM DATA, SOUTHWEST POWER STATION NO. 1

Unit rating	200 MW
Fuel characteristics	Coal, 3.5 percent sulfur
FGD vendor	Universal Oil Products
Process	Limestone scrubbing
New or retrofit	New
Start-up date	April 1977
FGD modules	Two
Efficiency, particulates	99.7 percent
Efficiency, SO_2	80 percent
Sludge disposal	Stabilized sludge is disposed of in unlined pond

WIDOWS CREEK NO. 8

BACKGROUND INFORMATION

The Widows Creek Power Station of Tennessee Valley Authority (TVA) is located in Bridgeport, Alabama. Widows Creek No. 8 is now the only coal-fired boiler at the plant that is equipped with an FGD system; a scrubber is also planned for Unit 7. Unit 8 has a capacity of 550 MW and uses coal with a sulfur content of approximately 3.7 percent.

POLLUTION CONTROL

The FGD system was designed and built by TVA. It consists of four Polycon limestone scrubber modules downstream of an ESP. Each scrubber is designed to handle 25 percent of the total flue gas flow. The modules each consist of a large venturi throat to compensate for poor particulate collection of the installed ESP, a grid type absorber, a Chevron vane type demister and a steam reheater. The overall particulate and SO_2 removal efficiencies of the modules are 99.5 percent and 80 percent, respectively.

The FCD system is still in the debugging stage and has not yet achieved simultaneous running of all four modules. Shakedown is due to be completed at the end of December.

Sludge produced by the FGD system is deposited in an unstabilized form in an unlined pond. FGD system data are presented in Table B-8.

Data on SO_2 collection efficiency and operability are not yet available on this FGD system since it started operation only recently as shown in Table B-8.[17,18,19]

Table B-8. FGD SYSTEM DATA, WIDOWS CREEK No. 8

Unit rating	550 MW
Fuel characteristics	Coal, 3.7 percent sulfur
FGD vendor	Tennessee Valley Authority
Process	Limestone scrubbing
New or retrofit	Retrofit
Start-up date	May 1977
Efficiency, particulates	99.5 percent
Efficiency, SO_2	80 percent
Sludge disposal	Unstabilized sludge is disposed of in unlined pond

SOUTH CAROLINA PUBLIC SERVICE CO.
WINYAH NO. 2

BACKGROUND INFORMATION

The Winyah Power Station of South Carolina Public Service Company is located in Georgetown, South Carolina. The plant now operates two coal-fired boilers; however, only Boiler 2 is equipped with an FGD system. Unit 2 has a rating of 280 MW, which is 35 MW less than Unit 1. Both boilers use a low-sulfur coal(approximately 1 percent sulfur) as fuel.

POLLUTION CONTROL

Two Babcock and Wilcox limestone scrubbing modules are located downstream of an ESP at Winyah Unit 2. The two new modules scrub only 50 percent of the total flue gas and remove 69 or 70 percent of the SO_2 in this stream. The particulate removal efficiency of the system is 99.4 percent.

Since the system started up in July 1977, the only reported problems are those with electrical relays. Sludge produced by the FGD system is deposited in an unstabilized, unlined pond, which is estimated to have a 20-year capacity. FGD system data are presented in Table B-9.

FUTURE PLANS

South Carolina Public Service intends to build two more boilers at the Winyah site. These will be of the same size as the Boiler 2 (280 MW).

Table B-9. FGD SYSTEM DATA, WINYAH NO. 2

Unit rating	280 MW
Fuel characteristics	Coal, 1 percent sulfur
FGD vendor	Babcock & Wilcox
Process	Limestone scrubbing
New or retrofit	New
Start-up date	July 1977
FGD modules	Two
Efficiency, particulates	99.4 percent
Efficiency, SO_2	70 percent
Sludge disposal	Unstabilized sludge is disposed of in unlined pond

PERFORMANCE HISTORY

Figures on availability, reliability, and operability are not yet available on the FGD system because of its very recent start-up.[20,21,22,23]

WILL COUNTY NO. 1

BACKGROUND INFORMATION

The Will County Power Station of Commonwealth Edison Company is located near Romeoville, Illinois. The plant has four electric generating units with a total rated capacity of 1147 MW. Only Unit 1 has been retrofitted with an FGD system.

Unit 1, a wet-bottom, pulverized-coal-fired boiler designed by Babcock & Wilcox, was installed in 1955. Rated at 167 MW, the unit now burns a low-sulfur (0.4%) Western coal and a high-sulfur (4.0%) coal from Illinois.

POLLUTION CONTROL

The boiler is fitted with an ESP manufactured by Western Precipitation Division. The ESP has a 79 percent actual particulate collection efficiency, and it is generally used when the FGD system is out of service. The present particulate emission rate from the FGD system is equivalent to 0.06 lb/MM Btu.

In 1972, the boiler was retrofitted with an FGD system for particulate and SO_2 control. Designed by Babcock & Wilcox, the wet limestone system removes 99.7 percent of the particulate matter and 82 to 90 percent of the SO_2. FGD system data are presented in Table B-10.

Table B-10. FGD SYSTEM DATA, WILL COUNTY NO. 1

Unit location	Romeoville, Illinois
Unit rating	167 MW
Fuel characteristics	Coal, 4.0 percent sulfur (design)
FGD vendor	Babcock & Wilcox
Process	Limestone scrubbing
New or retrofit	Retrofit
Start-up date	February 1972
Efficiency, particulates	98 percent
Efficiency, SO_2	82 percent
Water makeup	Open-loop, 1.66 gpm/MW
Sludge disposal	Stabilized sludge is disposed of in lined pond

The two scrubber modules were designed for 385,000 acfm at 355°F and handle 50 percent of the total boiler exhaust gas flow. Flue gas passes through the existing ESP and enters the venturi scrubber. Particulate removal efficiency is maintained by regulating the pressure drop across the adjustable venturi throat to maintain a pressure drop of about 9 inches of water. Gas velocity through the venturi is about 135 ft/sec.

The quenched flue gas flows upward through the SO_2 absorber tower and passes through two perforated trays. The trays are wetted with limestone slurry sprays located above the trays. The trays provide an extended wetted surface for absorption of SO_2 by the circulated slurry. Superficial velocity through the absorber is 12.2 ft/sec. Pressure drop through the two trays is 6 inches of water. Total system pressure drop is 25 inches of water.

The cleaned flue gas passes upward through a two-stage Z-shaped demister and then enters the reheater unit, where its temperature is raised from 128° to about 165°F. Reheat is necessary to prevent condensation in the fans, ducts and the existing brick-lined stack and to increase plume buoyancy.

Shortly after start-up in 1972 and during the initial shakedown, the plant encountered many problems. Demister plugging was an early problem and the demisters were eventually replaced. Modifications to the reheater were effected to reduce a vibration problem. Erosion and corrosion of equipment were also problems. Commonwealth Edison shut down Module B to concentrate on solving the problems of Module A, and Module B was not restarted until May 1975. Module A was then taken out of service in June 1975 to allow installation of new demisters and a new reheater. Because of delivery delays on the new reheater for Module A, this module was not restarted until late March 1976. Additional scrubber outages were due to delays in receiving replacement parts and a lack of sludge disposal pond capacity.

PERFORMANCE HISTORY

FGD system operability for this system is highly variable, but has been generally poor. The FGD system was designed for 76 percent SO_2 removal, and the utility reports an actual removal efficiency of 82 percent.

A summary of the reasons for recent poor operability is:[24,25,26]

March 1976 - The B-side module remained out of service until March 29 because of delays in the installation of a repaired reheater. During March, high-sulfur Illinois coal was fired in the boiler.

July 1976 - The A-side remained out of service the entire month because of repairs to a damaged absorber pump and pond thickener sludge overloading.

Oct.-Nov. 1976 - Problems related to the A-side were due primarily to repairs to a recycle pump and delays in the shipment of replacement parts.

May 1977 - The inlet valve on the recycle pump failed and the pond water return pump failed.

Jan. 1977 - The spent slurry valve failure, which occurred in December, kept the B-side out of service until January 27.

April 1977 - B-side outages occurred twice for absorber suction head repair and once to wash the I.D. booster fan.

May 1977 - Both modules were out of service for the first 15 days of the month due to transformer failure on the pond-water return pump. In addition, side A was out once because of inlet valve failure on an absorber pump.

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